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16. ABSTRACT

The two-volume proceedings document presentations at EPA's Sixth Symposium on Flue Gas Desulfurization (FGD), October 28-31, 1980, in Houston, Texas. Presentations covered such subjects as approaches for control of acid rain, the Nation's energy future, economics of FGD, legislative/regulatory developments, FGD research/development trends, FGD system operating experience, FGD byproduct disposal/utilization, developments in dry FGD, and industrial boiler applications.

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PREFACE

These proceedings for the symposium on "Flue Gas Desulfurization" constitute the final report submitted to the Industrial Environmental Research Laboratory, U.S. Environmental Protection Agency (IERL-EPA), Research Triangle Park, NC. The symposium was conducted at the Shamrock Hilton Hotel in Houston, TX, October 28-31, 1980.

This symposium was designed to provide a forum for the exchange of information, including recent technological and regulatory developments, on the application of FGD to utility and industrial boilers. The program included a Keynote Address on the approaches for control of acid rain, forecasts of energy and environmental technologies and economics for the 1980's, and sessions on the impact of recent legislation and regulations, research and development plans, utility applications, by-product utilization, dry scrubbing and industrial applications. Participants represented electric utilities, equipment and process suppliers, state environmental agencies, coal and petroleum suppliers, EPA and other Federal agencies.

Michael A. Maxwell, Chief, Emissions/Effluent Technology Branch, Utilities and Industrial Power Division, IERL-EPA, Research Triangle Park, NC, was General Chairman, and Julian W. Jones, a Senior Chemical Engineer in the same branch was Project Officer and Co-Chairman.

Franklin A. Ayer, Manager, Technology and Resource Management Department, Center for Technology Applications, Research Triangle Institute, Research Triangle Park, NC, was symposium coordinator and compiler of the proceedings.

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Jerome Rossoff, Chairman
The Aerospace Corporation
Los Angeles, California

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INTRODUCTION TO BY-PRODUCT UTILIZATION SESSION

BY JEROME ROSSOFF, CHAIRMAN

Our subject this morning is Flue Gas Cleaning By-Product Disposal and Utilization. With the passage of RCRA and the subsequent issuance of regulations by EPA concerning hazardous and nonhazardous wastes, one might ask, "What is the status of coal-ash and FGD wastes in light of the new regulations?" Briefly stated, it is as follows:

These wastes are temporarily exempted from regulation as hazardous wastes under EPA's hazardous waste management system until EPA conducts detailed, comprehensive studies of the environmental effects of coal ash and FGD waste disposal. Some of those studies will be discussed in this session.

Furthermore, federal criteria for disposal of nonhazardous wastes have been proposed by EPA for use by the states. While these criteria apply to all nonhazardous wastes, EPA intends ultimately to issue specific guidance for disposal of coal ash and FGD wastes. This guidance will be developed from the comprehensive studies just mentioned.

Therefore, disposal techniques are still of primary importance to effect land reclamation, and to prevent disposal site runoff and seepage from carrying excessive concentrations of suspended solids and dissolved salts to water supplies and the food chain.

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As identified on these charts, on-going surveys by EPA, EEI and the National Ash Association show that the current annual production of ash and FGD By-Products is approximately 75 million metric tons (on a dry basis) and by the year 2000 it is expected to be about 175 million metric tons (dry), with the FGD portion being about 30 percent in the year 2000. Since up to one-fourth of all power plant ash in this country is used in commercial applications, the total material to be disposed of in the year 2000 will be about 140 million tons. Considering total quantities of material to be handled, for example, if we assume the moisture content to be, say about 30 percent on the average, the actual quantity of combined products for disposal in the year 2000 would be about 200 million metric tons. Not only is there an appreciable cost involved, but this identifies an extremely large volume of material remaining to be handled and managed for disposal or possible utilization.

At past FGD symposia, we have heard many papers that discussed waste disposal programs involving mostly laboratory or small scale field evaluations of ponding and landfilling, conceptual design studies, preliminary assessments, and plans for commercial systems. These efforts, which have paved the way for operational procedures, included such items as: the physical and chemical characteristics of the waste materials, and evaluations of leachate, runoff, chemical treatment, oxidation, dewatering, handling, site management and disposal costs. We will hear more data on some of these factors today, but, in general the industry has reached a point where getting the job done is more apparent

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FGD WASTE PRODUCTION

	1980	1985	2000
● COAL BURNING UTILITY CAPACITY (MW)	235,000	301,000	498,000
● CAPACITY FACTOR (%)	48	57	65
● FGD SCRUBBING CAPACITY (MWeq)	27,000	61,000	163,000
● PERCENT NONREGENERABLE	93	87	83
● UTILITY AND INDUSTRIAL FGD WASTE WITHOUT ASH (10^6 metric tons, dry)	7	17	51
● TOTAL UTILITY AND INDUSTRIAL FGD WASTE + ASH (10^6 metric tons, dry)	74	97	174

Production and Utilization of Coal Ash

	1980	1985	2000
• COAL ASH PRODUCTION, UTILITY AND INDUSTRIAL (10^6 metric tons, dry)	67	80	123
• COAL ASH UTILIZATION (10^6 metric tons, dry)	14	18	29
• PERCENT UTILIZATION	21	22	24

than studying how to do it. This is not to say there is no room for learning or for improvement, which will be obvious in today's papers, however, the emphasis now is on the following: disposal applications in the field, utilization feasibility, and regulatory development.

In this regard, there are six highly pertinent papers for presentation today. These will provide: (1) a status report on the EPA programs which will be monitoring 12 full-scale utility waste disposal sites for the purpose of determining the degree to which these wastes must be managed for environmental protection; (2) a Department of Energy status report on the economic impact of RCRA on the electric utility industry; (3) an EPRI report on selected full-scale FGD sludge disposal demonstrations and site monitoring for environmental impact; (4) a report on a combined EPA/DOE project to study the disposal of FGD wastes in lignite surface mine pits; (5) a report on a program sponsored by EPA, EPRI, DOE, and two New York State agencies, which is demonstrating the use of chemically stabilized FGD wastes to construct artifical reefs; and (6) an EPA report on federal procurement guidelines for the use of fly ash in cement and concrete production. These subjects highlight the fact that we are moving rapidly from bench scale and prototype studies to operational applications.

CHARACTERIZATION AND ENVIRONMENTAL MONITORING OF FULL-SCALE UTILITY WASTE DISPOSAL:

A STATUS REPORT

Chakra J. Santhanam
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Cambridge, MA 02140

and

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U.S. Environmental Protection Agency
Industrial Environmental
Research Laboratory
Research Triangle Park, NC 27711

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ABSTRACT

This paper presents a status report on the EPA project entitled "Characterization and Environmental Monitoring of Full-Scale Utility Waste Disposal Sites" (EPA Contract No. 68-02-3167). Arthur D. Little, Inc. is the prime contractor on this project which involves the characterization, environmental monitoring, and engineering/economic assessment of coal ash and flue gas desulfurization waste disposal at 12 full-scale waste disposal sites. The project is designed to obtain technical background data and information so that EPA can determine the degree to which disposal of these wastes (from coal-fired power plants) needs to be managed in order to protect human health and the environment. This effort will fulfill some of EPA's responsibilities under the Resource Conservation and Recovery Act.

To date, the major accomplishments of this project include:

- Evaluation of available data on coal-fired power plants in the United States to develop a list of candidate and backup sites,
- Preparation of procedures manuals, and
- Progress in securing utility involvement and cooperation.

ACKNOWLEDGEMENTS

The authors gratefully acknowledge the assistance of the following members of the Arthur D. Little, Inc. staff in the preparation of this paper: Anne B. Littlefield, Armand A. Balasco, David E. Kleinschmidt, and Charles B. Cooper.

1.0 INTRODUCTION

The U.S. Environmental Protection Agency's (EPA) Office of Research and Development (ORD) has contracted with Arthur D. Little, Inc. (ADL) to conduct a study of current coal ash and flue gas desulfurization (FGD) waste disposal practices at coal-fired power plants (EPA Contract No. 68-02-3167). EPA's Office of Solid Waste (OSW) is working closely with ORD on this program. The study involves environmental monitoring; data gathering; and engineering, environmental, and economic assessments of disposal practices (for coal ash and FGD wastes) at 12 full-scale waste disposal sites at various locations around the country. The study results will provide the technical background data and information needed to determine the degree to which disposal of these wastes needs to be managed in order to protect human health and the environment. This paper is intended to provide a status report on the project as of August 1980.

2.0 OVERVIEW OF THE PROJECT

2.1 Goals

The study will involve environmental monitoring, data gathering, and economic assessment of disposal practices at utility solid waste (coal ash and FGD waste) disposal sites at various locations around the country. Sites will be selected so that the results of the study will provide the technical background data and information needed to determine the degree to which disposal of these wastes needs to be managed in order to protect human health and the environment. It is anticipated that EPA will ultimately issue to Federal and state permitting officials guidelines for disposal of coal ash and FGD wastes (which together are called flue gas cleaning or FGC wastes) under the Resource Conservation and Recovery Act of 1976 (RCRA).

After establishing these goals for the project, EPA assessed two factors:

- the types of coal ash and FGD wastes sent to disposal and the methods of disposal employed, and
- the funding available for the project.

Consideration of these factors led to a decision by EPA to characterize and monitor 12 sites (each site to be monitored for about 1 year). Arthur D. Little, Inc. was selected as the prime contractor for the project (EPA Contract No. 68-02-3167); work on the project was initiated in October 1979.

2.2 Generation of Utility FGC Wastes

2.2.1 Technology and Production of Wastes

Coal-fired utility boilers generate two types of coal ash: fly ash and bottom ash. Fly ash is collected by mechanical collectors, electrostatic precipitators, fabric filters, or wet scrubbers. Flue gas desulfurization can be accomplished by nonregenerable or throwaway systems which result in FGD wastes and regenerable systems which primarily produce a saleable product (sulfur or sulfuric acid). At present approximately 27,000 Mw gross generating capacity (25,000 MW scrubbed capacity) of coal-fired utility boilers employ FGD systems (1); over 90% of these use nonregenerable systems. By July 1980, 95,000 MW of generating capacity was committed to FGD systems (in planning, contract awards, or under construction) (2). Nonregenerable systems currently employ wet scrubbing technology, although some dry scrubbing systems are expected to be operational before 1982. The principal types of systems used in utility power plants are those based on direct limestone, direct lime, alkaline fly ash, and dual alkali.

The quantities of FGC wastes depend on the ash and sulfur content of the coal, emission regulations, the types of ash collection and FGD systems, and operating conditions of the systems and boiler. To meet New Source Performance Standards, a typical utility operating at 70% load produces 100 to 500 metric tons (110 to 550 tons) of dry ash-free FGD sludge and 200 to 600 metric tons (220 to 660 tons) of coal ash annually per megawatt of capacity. Table 2.1 shows one estimate of present and anticipated FGC waste generation in the United States.

2.2.2 Number of Coal-Fired Utility Plants

There are currently more than 340 steam electric plants in the United States utilizing coal for 80% or more of their power generation and with nameplate generating capacity greater than 25 MW. By 1988 it is anticipated that an additional 271 coal-fired plants will be built with a total generating capacity of 142,000 MW (3). In addition, a number of plants are likely to be converted to coal from other fossil fuels (oil and gas). The development of baseline data under this project takes cognizance of current and anticipated use of coal in power plants.

2.2.3 Disposal vs. Utilization Options

Today most of the coal ash and all of the FGD wastes generated are sent to disposal. Considering the expected increase in coal consumption in boilers in the United States, this is likely to be the case for many years. Utilization of FGC wastes is expected to grow but at a slower rate than FGC waste generation. Over the longer term, an effective way to manage coal and FGD wastes is to utilize them. There is currently some utilization of coal ash but no utilization of FGD wastes in the United States.

Currently, all FGC wastes are disposed of on land. At-sea disposal may be a future alternative if it can be practiced under environmentally acceptable conditions. The principal methods of disposal are:

TABLE 2.1

PROJECTED GENERATION OF COAL ASH AND FGD WASTES

			PROJECTED			
	1975		1985		2000	
	10^3 Metric Tons	% of Total	10^3 Metric Tons	% of Total	10^3 Metric Tons	% of Total
Coal Ash						
Industrial	-	-	8,590	12	19,950	19
Utility	-	-	64,440	88	84,800	81
Total	52,060	-	73,030	100	104,750	100
FGD Wastes						
Industrial	-	-	1,090	5	5,260	15
Utility	-	-	21,050	95	29,860	85
Total	6,200	-	22,140	100	35,120	100

Source: Reference 2.

- ponding,

- landfilling, including mine disposal which may be considered a special subcategory of landfilling, and

- interim ponding followed by landfilling.

On a national basis ponding is a more widely employed disposal practice than any other today. Ponds can be designed based on diking or incision and can even be engineered on slopes, but the construction of dikes or other means of containment for ponds is usually expensive. In the past few years, landfilling has become a growing practice. Typically, in landfilling the wastes are dewatered to a moist soil-type material (unless the wastes are collected dry) and transported to the disposal site where they are spread on the ground in about 0.3-to-1-meter (1-to-3 foot) heights and compacted. Layering proceeds and ultimately fill may be about 8 to 25 meters (25 to 80 feet) or more in height. Interim ponding/landfilling is a hybrid involving ponding followed by landfilling of the dredged solids. This method combines the ease of handling of a pond with the limited land requirements of a landfill.

2.2.4 Future Trends

In the past, the majority of utilities operating FGC systems have typically disposed of wastes by storage in ponds. However, the following factors will increase the sources and total volume of waste and influence disposal options in the coming years:

- An increase in coal-fired capacity in the United States. In 1976 the total U.S. coal-fired electric-utility generating capacity was estimated at over 191,000 MW in 399 plants (4). The estimated coal-fired capacity is expected to increase by 1988 to over 330,000 MW (3).

- A major increase in the application of FGD scrubber technology by utilities and a consequent increase in FGD waste generation. At present, over 27,000 MW of generating capacity at 39 plants utilize FGD systems, and over 95,000 MW of gross generating capacity (∿ 91,000 MW scrubbed capacity) has been committed (i.e., under construction, contract awarded, or operating today) to FGD systems (6).

- "Stabilization" of FGD wastes, that is making them into moist soil-like material by processes that usually involve addition of lime and fly ash. Advances in such stabilization technology for FGD wastes have led to reduced cost and increased reliability of such technology and permit landfilling of partially dewatered solids instead of ponding of sludges that are difficult to handle. In the future, disposal of wastes in managed fills is likely to be encouraged.

- Environmental regulatory developments including the Clean Air Act of 1977 and RCRA. New Source Performance Standards for criteria pollutants were issued in June 1979 (5). RCRA regulations for hazardous waste disposal were issued in May 1980. Those regulations and recently passed Congressional amendments provide a temporary deferral for fly ash, bottom ash, boiler slag, and flue gas emission control wastes and exclude them from any RCRA hazardous waste regulations pending completion of this project and other studies. Other utility wastes which may be hazardous are subjected to existing RCRA hazardous waste regulations. On the other hand, guidelines to the states for management of nonhazardous waste (which would include utility solid wastes) under Sections 1008 and 4004 of RCRA were issued on September 13, 1979, and currently apply to fly ash, bottom ash, boiler slag, and flue gas emission control waste.

- Energy-related developments under the Power Plant and Industrial Fuel Act of 1978. Several plants are scheduled to switch to coal (6) from other fossil fuels (oil and gas).

2.3 Environmental Impacts and Regulatory Requirements

The environmental impacts of FGC waste disposal and hence any potential threat to human health and the environment are influenced by three factors:

- type of waste generated (physical and chemical characteristics),

- disposal method employed (ponding, landfilling, or other), and

- disposal site characteristics (soil type, hydrogeology, climate, etc.).

In this project, a mix of waste types, modes of disposal, and site characteristics will be subjected to characterization and environmental monitoring. The disposal methods examined will include the most prevalent methods used in the industry today, as well as those which are likely to represent the best control technology standards for disposal of coal ash alone, coal ash and FGD waste combined, and, if appropriate, FGD waste alone. Each specific disposal operation will be evaluated for its current and potential impact on ground/surface water quality and other environmental impacts in the vicinity of the disposal site.

Criteria for planning and implementation of this project have been prioritized to reflect the relative importance of various potential impacts under RCRA to this particular effort. Specifically, highest priority is given to three subject areas that are both characteristically important for utility solid waste disposal and principal regulatory responsibilities under RCRA. These are:

- groundwater quality,

- surface water quality from non-point sources, and

- use of potentially mitigative design, management, or control practices.

2.4 Scope and Approach

2.4.1 Overall Scope

The project consists of three major tasks:

- Task I - Site Selection and Preliminary Test Plan Preparation,
- Task II - Site Development, Characterization, and Environmental Monitoring, and
- Task III - Environmental and Engineering/Economic Assessment

The final report at the end of Task III will include recommendations on site criteria, control technology options, operating practices, monitoring, and economics (capital and operating costs) of FGC waste disposal.

2.4.2 Task I

This initial task consists of:

- Evaluation of available data to select 12 suitable sites. The basic approach here has been to develop a pool of candidate and backup sites for detailed evaluation and field visits. The final 12 sites will be selected as a result of this effort.
- Definition of baseline procedures and approaches for development of an FGC waste disposal site, characterization of the waste and surrounding soils, environmental monitoring, and engineering/economic data gathering.
- Preparation of preliminary test plans on the 12 sites. The test plans, after approval by EPA, will become the basis for work at each site.
- Analysis of samples by RCRA Section 3001 protocol. During the site selection process, a number of site visits will be made. During these initial visits samples of FGC wastes will be collected and tested according to the RCRA Section 3001 protocol, which includes an extraction procedure (EP). The protocol specifies that if the concentration of any of eight trace metals (arsenic, barium, cadmium, chromium, lead, mercury, selenium, and silver) in the EP extract of a waste exceeds 100 times the Federal Drinking Water Standards, the waste is considered toxic, and therefore is "hazardous". An analysis of available data supports the view that the maximum concentrations of these trace metals in the extract from FGC wastes will probably be between 4 and 20 times the Drinking Water Standards. Unless the results of the

analyses (conducted under this study) indicate otherwise, these efforts are not expected to have a major effect on the site selection process.

The information flow diagram for Task I in Figure 2.1 shows the overall work structure for the preparation of preliminary site recommendations and test plans. Work in Task I is envisioned as an iterative process involving screening of potential sites, development of preliminary plans for testing at each of the sites selected, and estimation of costs for testing at each site. The intent of the iterative procedure is to ensure, to the extent allowed by this preliminary analysis, the recommendation of the most cost-effective program capable of obtaining the data needed for support of RCRA.

The recommended set of sites, preliminary test plans, and estimated costs will then be submitted to EPA for review and approval. This review may necessitate reevaluation of potential alternative sites and/or adjustment of preliminary test plans and costs. Such a reevaluation and adjustment could result from consideration of a number of factors, including anticipated total testing costs, which could lead to some restructuring of the matrix of testing sites.

The approach used in developing the plans for characterizing and sampling a disposal site will be aimed at providing information on areas relevant to potential RCRA environmental guidelines. On a site-specific basis, this will include:

- Waste Characterization
  - Characterization of the previously and currently disposed of wastes by detailed chemical analysis.
  - Characterization of waste materials using physical and engineering tests to assess the possibility of site reclamation.
- Site Characterization
  - Characterization of the groundwater around the site and the underlying soil in the unsaturated infiltration zone to determine the extent and chemical composition of leachate under the site.
  - Characterization of runoff from the site with the intent of assessing the extent of potential environmental impact via this mechanism.

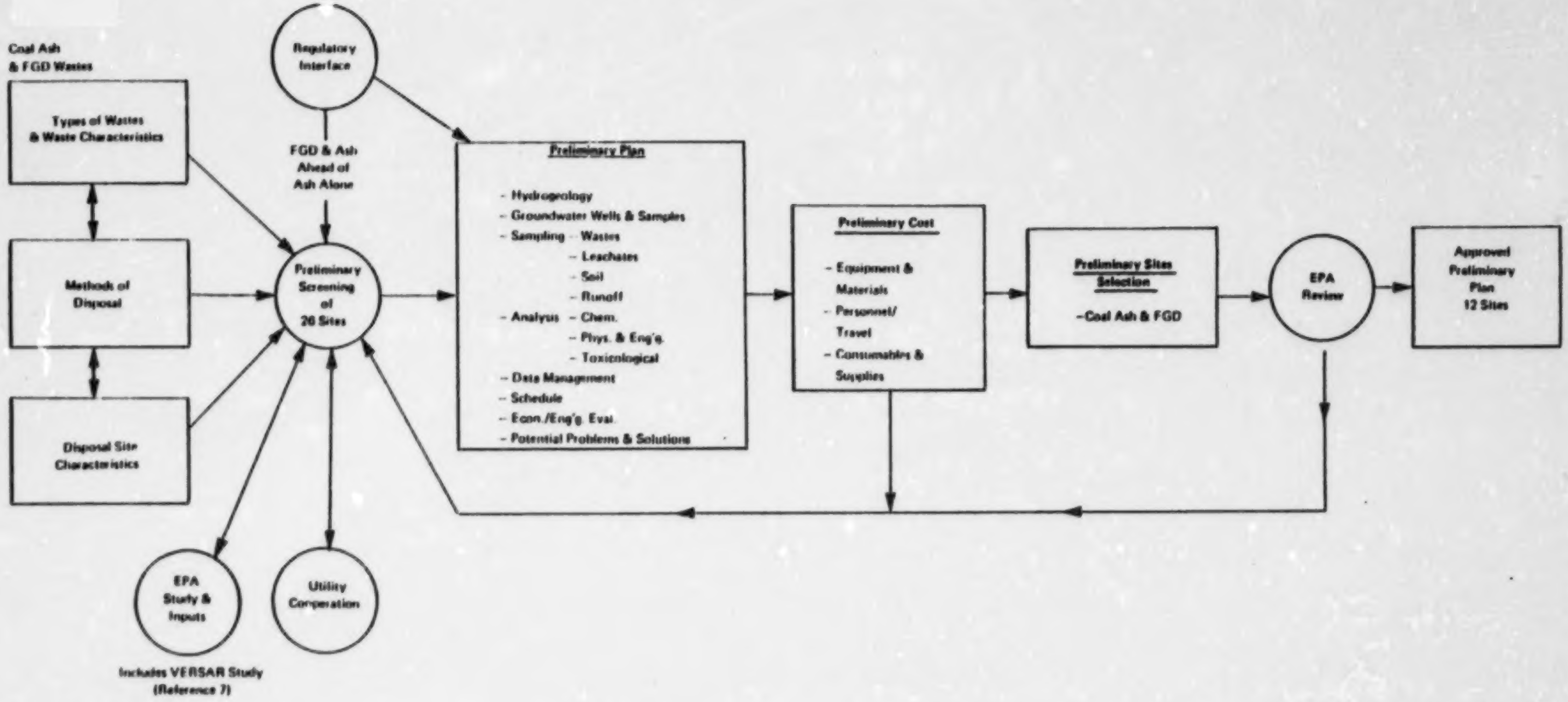
Coal Ash
& FGD Wastes
Types of Wastes & Waste Characteristics
Methods of Disposal
Disposal Site Characteristics
Regulatory Interface
FGD & Ash Ahead of Ash Alone
Preliminary Screening of 20 Sites
EPA Study & Inputs
Includes VERSAR Study (Reference 7)
Utility Cooperation
Preliminary Plan
- Hydrogeology
- Groundwater Wells & Samples
- Sampling - Wastes
- Leachates
- Soil
- Runoff
- Analysis - Chem.
- Phys. & Eng'g.
- Toxicological
- Data Management
- Schedule
- Econ./Eng'g. Eval.
- Potential Problems & Solutions
Preliminary Cost
- Equipment & Materials
- Personnel/ Travel
- Consumables & Supplies
Preliminary Sites Selection
-Coal Ash & FGD
EPA Review
Approved Preliminary Plan 12 Sites

FIGURE 2.1 TASK I INFORMATION FLOW DIAGRAM

- Characterization of air-suspended materials and determination of volatile species which may lead to environmental impact via these mechanisms.

2.4.3 Task II

Work in Task II consists basically of two efforts--first, the finalization of test plans for each site, including revised cost estimates, where appropriate; and second, disposal site and waste characterization, environmental monitoring, and the obtaining of capital and operating cost information on FGC waste disposal. The formulation of test plans will begin as site selection proceeds during Task I. It is anticipated that the actual process for arriving at the final set of sites may be iterative to some extent, involving interface with the utilities to achieve their cooperation, review of recommended sites and alternatives, and preparation of test plans. Figure 2.2 presents the information flow under Task II.

It is anticipated that finalization of test plans, including, if appropriate, revision of Task I preliminary cost estimates, and review and approval of the plans by EPA, may also require some iterative refinements. Substantial variations in revised cost estimates from the preliminary Task I estimates for a site may require reconsideration of the specific site or even the overall structure of the site matrix.

The basis for the preliminary test plans and their finalization under Task II will be an evaluation of existing information and data relating principally to:

- Extent of utility cooperation and any special requirements established by a utility;
- Waste properties and variations in waste properties;
- Disposal site operating history and current practices;
- Site characteristics, including hydrogeology, geotechnical characteristics, climate, and surrounding land use;
- Existence and availability of data from prior monitoring and analytical studies of the waste, waste disposal area, and surrounding region;
- Existence and availability of groundwater wells (on- and off-site) and other monitoring stations (e.g., meteorological) for use in the study; and
- Existence, extent, and availability of economic data.

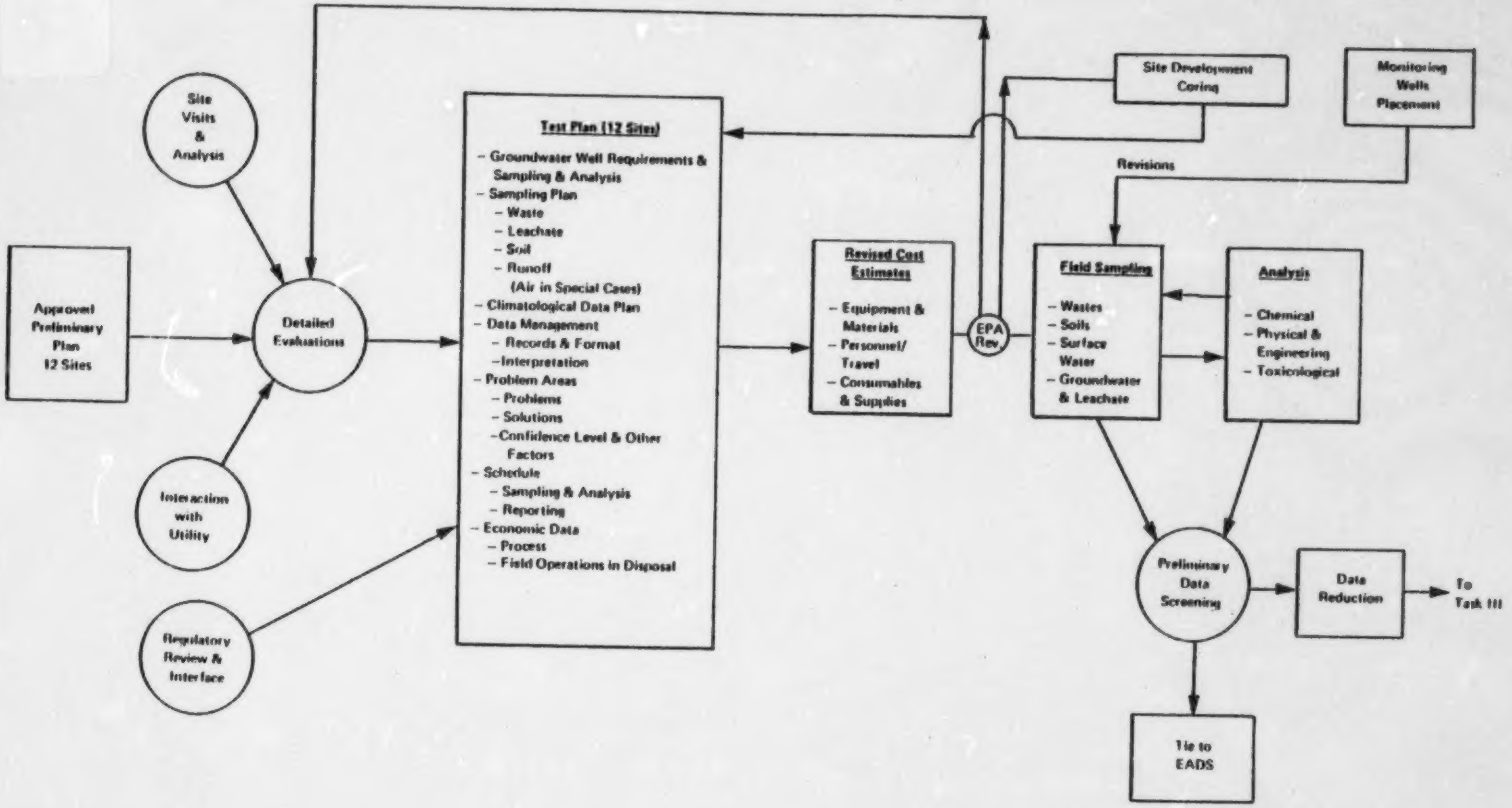
Approved Preliminary Plan 12 Sites
Site Visits & Analysis
Detailed Evaluations
Interaction with Utility
Regulatory Review & Interface
Test Plan (12 Sites)
– Groundwater Well Requirements & Sampling & Analysis
– Sampling Plan
– Waste
– Leachate
– Soil
– Runoff (Air in Special Cases)
– Climatological Data Plan
– Data Management
– Records & Format
–Interpretation
– Problem Areas
– Problems
– Solutions
–Confidence Level & Other Factors
– Schedule
– Sampling & Analysis
– Reporting
– Economic Data
– Process
– Field Operations in Disposal
Revised Cost Estimates
– Equipment & Materials
– Personnel/ Travel
– Consumables & Supplies
EPA Rev.
Site Development Coring
Monitoring Wells Placement
Revisions
Field Sampling
– Wastes
– Soils
– Surface Water
– Groundwater & Leachate
Analysis
– Chemical
– Physical & Engineering
– Toxicological
Preliminary Data Screening
Data Reduction
To Task III
Tie to EADS

FIGURE 2.2 TASK II INFORMATION FLOW DIAGRAM

It is anticipated that this information and these data will be derived from a number of sources, including visits to each utility and site selected, contacts with regional EPA and other Federal and state environmental agencies, and contacts with other organizations which have performed relevant site or regional studies. These visits and contacts will be initiated as soon as approval by EPA of the preliminary recommendations on candidate sites has been received.

As final site selections are made and test plans are approved, site development, waste characterization, and environmental monitoring will begin. In order to meet the required project schedule, site monitoring will be initiated in a staged fashion in concert with finalization of the selection of sites. In some cases, to increase the validity of the results, it may be desirable to monitor a particular site for an additional few months, depending on the site characteristics, plant operations, and the order of site evaluation startup. This can be accommodated within the existing project structure to the extent allowed by budget and monitoring requirements at other sites.

The monitoring effort will consist of sampling runoff, surface water, groundwater, and leachate for a 1-year period. Groundwater and leachate may be sampled monthly (on an average) while runoff and surface water will be sampled as appropriate. An adequate number of samples of wastes and surrounding soils will also be obtained during this period. The field sampling and analysis for all sites are currently anticipated to be completed by 1982.

2.4.4 Task III

The objective of this task is identical to that of the overall effort; namely, to develop the technical bases for guidelines that may be applied to the disposal of coal ash and FGD wastes under RCRA. This task is the one under which that objective will be achieved, based on the planning and data collection accomplished in Tasks I and II.

The proposed approach to executing this task reflects the incremental importance of each set of RCRA regulations to the various disposal considerations discussed below. Specifically, the data from Task II will be assessed in three categories:

- environmental,
- disposal operations, and
- disposal economics.

An information flow diagram outlining the approach to this task is shown in Figure 2.3. Major work under Task III, which involves the environmental, engineering, and economic assessments and the formulation

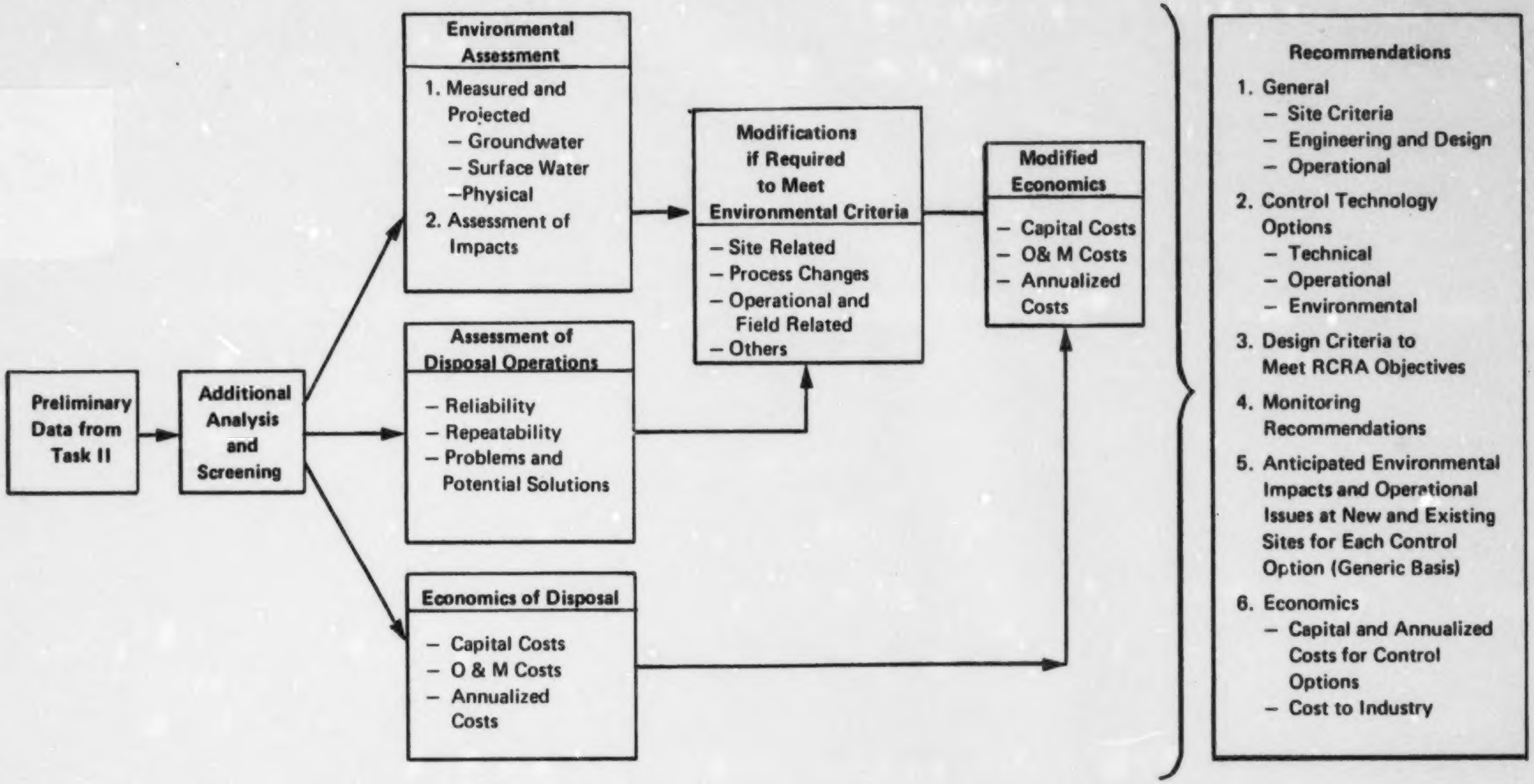
Preliminary Data from Task II
Additional Analysis and Screening
Environmental Assessment
1. Measured and Projected
– Groundwater
– Surface Water
–Physical
2. Assessment of Impacts
Assessment of Disposal Operations
– Reliability
– Repeatability
– Problems and Potential Solutions
Economics of Disposal
– Capital Costs
– O & M Costs
– Annualized Costs
Modifications if Required to Meet Environmental Criteria
– Site Related
– Process Changes
– Operational and Field Related
– Others
Modified Economics
– Capital Costs
– O& M Costs
– Annualized Costs
Recommendations
1. General
– Site Criteria
– Engineering and Design
– Operational
2. Control Technology Options
– Technical
– Operational
– Environmental
3. Design Criteria to Meet RCRA Objectives
4. Monitoring Recommendations
5. Anticipated Environmental Impacts and Operational Issues at New and Existing Sites for Each Control Option (Generic Basis)
6. Economics
– Capital and Annualized Costs for Control Options
– Cost to Industry

FIGURE 2.3 TASK III INFORMATION FLOW DIAGRAM

of recommendations, will be initiated part way through the completion of Task II, as site evaluations proceed. The work under Task III is scheduled to continue for about 3 months beyond the completion of field sampling and analysis under Task II.

All assessments will provide input for the formulation of a series of recommendations concerning the bases for the design and operation of FGC waste disposal facilities. While these recommendations will certainly reflect the knowledge gained from the studies at each site, recommendations will be made which are as broadly applicable as possible.

2.5 Organizations Involved

2.5.1 Contractor and Subcontractors

Arthur D. Little, Inc. (ADL) is the prime contractor on the project. Five subcontractors (Bowser-Morner Testing Laboratories, Inc., Haley & Aldrich, Inc., Kaiser Engineers Power Corp., TRW, Inc., and the University of Louisville) are assisting ADL. Table 2.2 shows the participants and their areas of responsibility.

2.5.2 Utility Interfacing

Coordination with the utilities and utility cooperation are key elements in the implementation of this project if it is to successfully achieve EPA's overall objectives of supporting RCRA. The requirements for FGC systems, in general, and regulation of waste disposal, in particular, are sensitive issues to many utilities. Two factors with respect to this project are expected to be of prime concern from the utilities' standpoint: (1) test work that indicates adverse environmental impacts for any specific site operation; and (2) the obvious reluctance to assist in the development of guidelines which may ultimately impose restrictions (and higher costs) on utilities. Utilities, however, have a strong interest in ensuring that necessary environmental protection measures are cost effective and technically sound. Hence, if they understand and agree with the objectives and approach of a program of this type, they are more likely to participate. However, they must also be assured that the compliance status of the study sites will not be jeopardized by the program. It is clear, therefore, that specific attention must be paid to establishing proper communication with the industry as a whole, as well as with specific utilities having waste disposal sites of interest.

To establish this communication, the following steps have been taken:

- General Communication - From the outset of the project, establishment of the best lines of communication with specific utilities having disposal sites of potential interest and with the entire

TABLE 2.2

CHARACTERIZATION AND ENVIRONMENTAL MONITORING OF FULL-SCALE UTILITY WASTE DISPOSAL SITES

MAJOR PARTICIPANTS

Program Area	Principal Participants
Prime Contractor/Project Management	Arthur D. Little, Inc. Cambridge, MA 02140
Chemical Sampling and Analysis	Arthur D. Little, Inc. Cambridge, MA 02140 and TRW, Inc. Redondo Beach, CA 90278
Engineering/Economic Evaluation	Arthur D. Little, Inc. Cambridge, MA 02140 and Kaiser Engineers Power Corp. Oakland, CA 94623
Geotechnical & Field Drilling	Bowser-Morner Testing Laboratories, Inc. Dayton, OH 45401
Hydrogeologic Activities	Haley & Aldrich, Inc. Cambridge, MA 02142
Physical Sampling & Analysis	University of Louisville Louisville. KY 45208 and Bowser-Morner Testing Laboratories, Inc. Dayton, OH 45401
Quality Assurance/Quality Control	Arthur D. Little, Inc. Cambridge, MA 02140

utility industry has been sought. The utilities will be given every opportunity to fully understand the objectives and nature of the project and it is hoped that they will realize that through their cooperation realistic results can be achieved. This will involve initial and followup contacts with specific utilities, as well as with appropriate Electric Power Research Institute (EPRI) and Edison Electric Institute (EEI) staff members who are familiar with waste disposal problems and utility concerns. These lines of communication need to be maintained and improved throughout the project.

- Review Committees - Two committees have been formed: the Advisory Committee with representatives of EPA, Department of Energy (DOE), EEI, and American Public Power Association (APPA); and the Technology Committee consisting of representatives of EPA, ADL, EPRI, and a number of utilities. These committees will review and comment on the project as it proceeds and serve as a mechanism for the exchange of ideas, concerns, and, where appropriate, data. Table 2.3 lists current member organizations of the committees.

- Data Review - Data and information obtained during the test program and the results of the work should be provided to the utilities involved for their review and comment prior to any formal release. This would give them a type of participation in the program that would help avoid any inadvertent misinterpretation of data.

2.6 Schedule

The original schedule estimate was that the project could be completed in 26 months (by November 1981). However, it is currently estimated that the project will be completed by November 1982. This "stretching out" of the schedule is due to at least three factors:

- First, the data base on utility site characteristics and waste disposal practices available at the beginning of the project was inadequate and contained a significant number of inaccuracies. Thus, it has been necessary for ADL to begin the site selection process with a complete review of all coal-fired power plants in the United States. This has been a much more comprehensive and time-consuming effort than was originally anticipated.

- Second, a diversion of some effort to a review with EPA of the overall project scope and direction was required. In the early months of the project, EPA was developing the final regulations for hazardous waste management under RCRA. The structure of this project was evaluated for some time during the final regulatory development effort.

TABLE 2.3

ORGANIZATIONS ON THE COMMITTEES

Advisory Committee (AC)	Technology Committee (TC)*
American Public Power Association	American Electric Power
Arthur D. Little, Inc.	Arthur D. Little, Inc.
Department of Energy	Duke Power
Edison Electric Institute	Electric Power Research Institute
Environmental Protection Agency	Environmental Protection Agency
	Los Angeles Department of Water & Power
	Northern States Power
	Southern Company Services
	Tennessee Valley Authority
	Union Electric

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*Other utilities will be invited to join the TC as the project proceeds.

- Third, although the strong participation by the Advisory and the Technology Committees in critically reviewing all major documentation has been beneficial to the project, this has also led to delays.

Figure 2.4 shows the present schedule for this project.

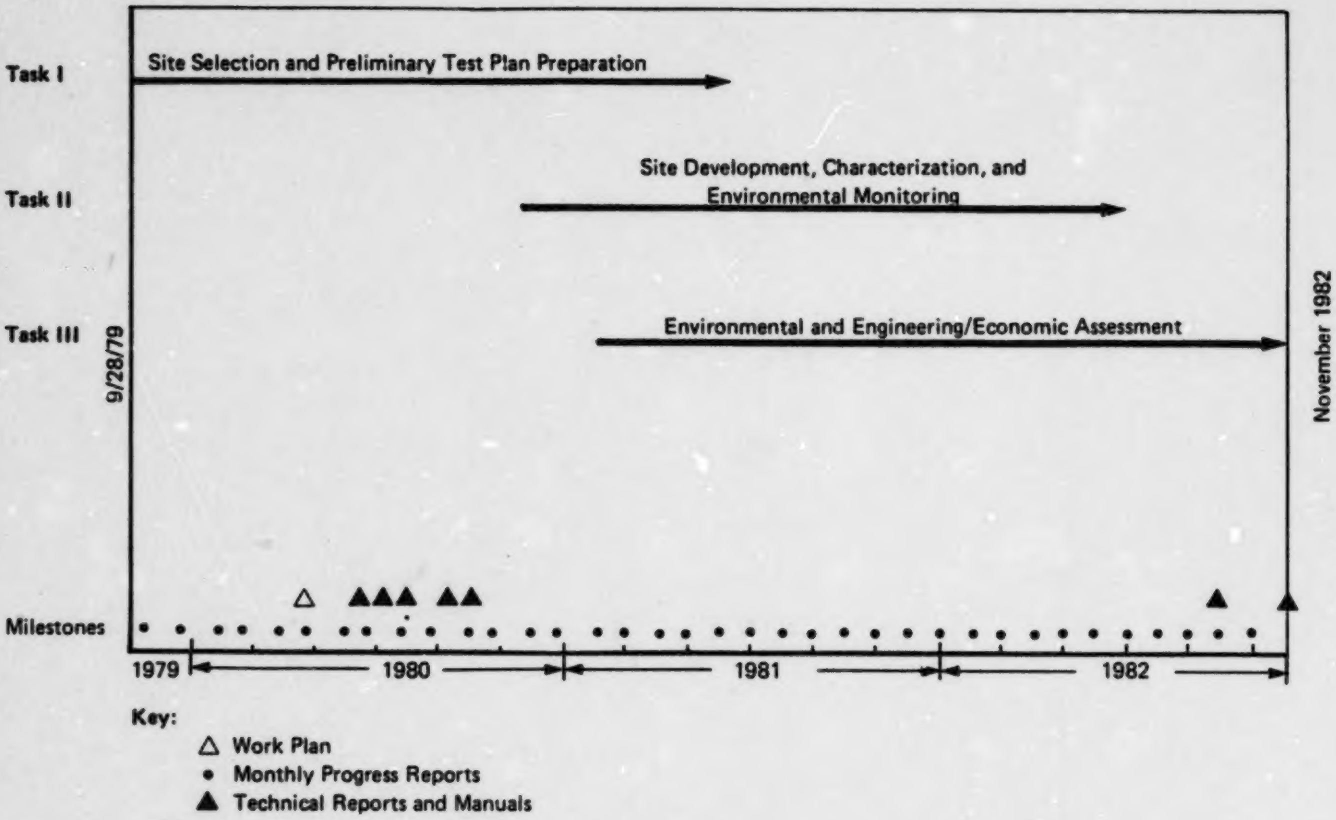


FIGURE 2.4 PROJECT SCHEDULE

3.0 ACCOMPLISHMENTS TO DATE (October 1979 to August 1980)

3.1 Overview

Major accomplishments on the project to date are:

- The evaluation of available data through detailed screening has led to the recommendation of a pool of 18 candidate and 8 backup sites for further evaluation which, in turn, will lead to the selection of the final 12 sites. A draft report entitled "Candidate Site Selection Report" has been submitted (8).
- The preparation of four procedures manuals (see Section 3.3, below) detailing conduct of various items of work under this project is underway. Two of these, "Hydrogeologic and Geotechnical Procedures Manual" and "Sampling and Analysis Procedures Manual," have been completed in draft form and are being reviewed.
- Substantial progress has been made in securing utility cooperation. Two committees have been set up and are working smoothly. The Advisory Committee and the Technology Committee are reviewing all major documentation on this project.
- Initial plant visits are underway; by the end of August 1980, 6 of 18 candidate sites had been visited.

Details on these accomplishments are described in the remainder of this section.

3.2 Site Selection Process

The overall approach in choosing 12 disposal sites for detailed characterization and environmental monitoring involves two steps:

- The first step was a preliminary evaluation of all coal-fired power plants in the United States on which data are available in order to select a number of candidate sites and a smaller number of backup sites.
- In the second step, the candidate sites will be subjected to closer evaluation, including site visits, preliminary testing of grab samples of wastes, and detailed geotechnical and hydrogeologic evaluation. On the basis of these evaluations, the final 12 sites will be selected.

The first step has been completed and a draft report has been submitted (8). The environmental impacts of FGC waste disposal, hence any potential threat to human health and the environment, are influenced by three factors:

- type of waste generated,
- disposal method employed, and
- disposal site characteristics (soil type, hydrogeology, climate, etc.).

Therefore, during the candidate site selection process, a mix of waste types, modes of disposal, and site characteristics was considered.

The selection of candidate and backup sites involved a two-step process:

- First, the contiguous 48 states were divided into 14 discussion regions and the plants in each region were screened to develop a list of plants suitable for consideration as candidate and backup sites. It was agreed that the total number of candidate sites would be 18, providing 50% redundancy. In addition, a smaller number of backup sites was considered desirable. A target of 25 to 30 sites, including backup sites, was established. Based on an assessment of present and future FGC waste disposal, a preliminary distribution of the targeted number of candidate sites in each region was agreed upon. During the screening process, the investigators remained cognizant of the targeted number in each region but were not absolutely limited by that number. The goal was to choose desirable plants in as many regions as possible. A list of 26 plants in all the regions came through this screening process.

- Second, these 26 plants were ranked during iterative group discussions, leading to the nomination of 18 as candidate and the remainder as backup sites.

3.2.1 Regional Division of the United States

In order to facilitate the site selection process, the United States was divided geographically into 14 discussion regions. This division resulted in a manageable number of plants for detailed and iterative evaluation by the investigators involved in the site selection process. Each discussion region corresponds to one or more physiographic regions.

Division of the country into these regions allowed the investigators to determine whether or not the disposal operations at a plant were typical of practices which were general throughout a region and the likelihood of continued or more extensive use of a paticular disposal practice. It also allowed the investigators to determine whether the hydrogeologic and environmental conditions at a site were representative of others in the region. (Assuming that these conditions were representative, the sites were also examined to determine that a reasonable assessment could be made in 1 year of monitoring.) The sites which appeared promising were compared with potential candidate sites in adjacent regions so that the most promising sites could be selected.

3.2.2 Targeted Number of Candidate Sites

Given that 12 sites will ultimately be chosen for monitoring, the investigators decided that the preliminary evaluations should result in the recommendation of 18 candidate sites and a smaller number of backup sites. The reasons for this decision were:

- While data are available, the investigators have not visited most of the sites. First-hand evaluations during field visits may reveal previously unknown problems leading to rejection of some of the 18 candidate sites.

- Issues relating to utility cooperation and other factors may lead to some sites being less suitable than others for this project.

There are currently more than 350 steam-electric plants in the United States with coal-firing capability. About 340 of these plants have greater than 25 MW capacity and utilize coal for more than 80% of their power production or generation. More than 90% of the capacity is in plants larger than 200 MW (9). Figure 3.1 presents data on the larger plants in each discussion region.

In order to arrive at a reasonable number of candidate sites for each region, a sensitivity to the actual number of coal-fired power plants in each region was required. Further considerations included:

- The existence of several types of waste and disposal mode characteristics in one region might require selection of more candidate sites in that region.

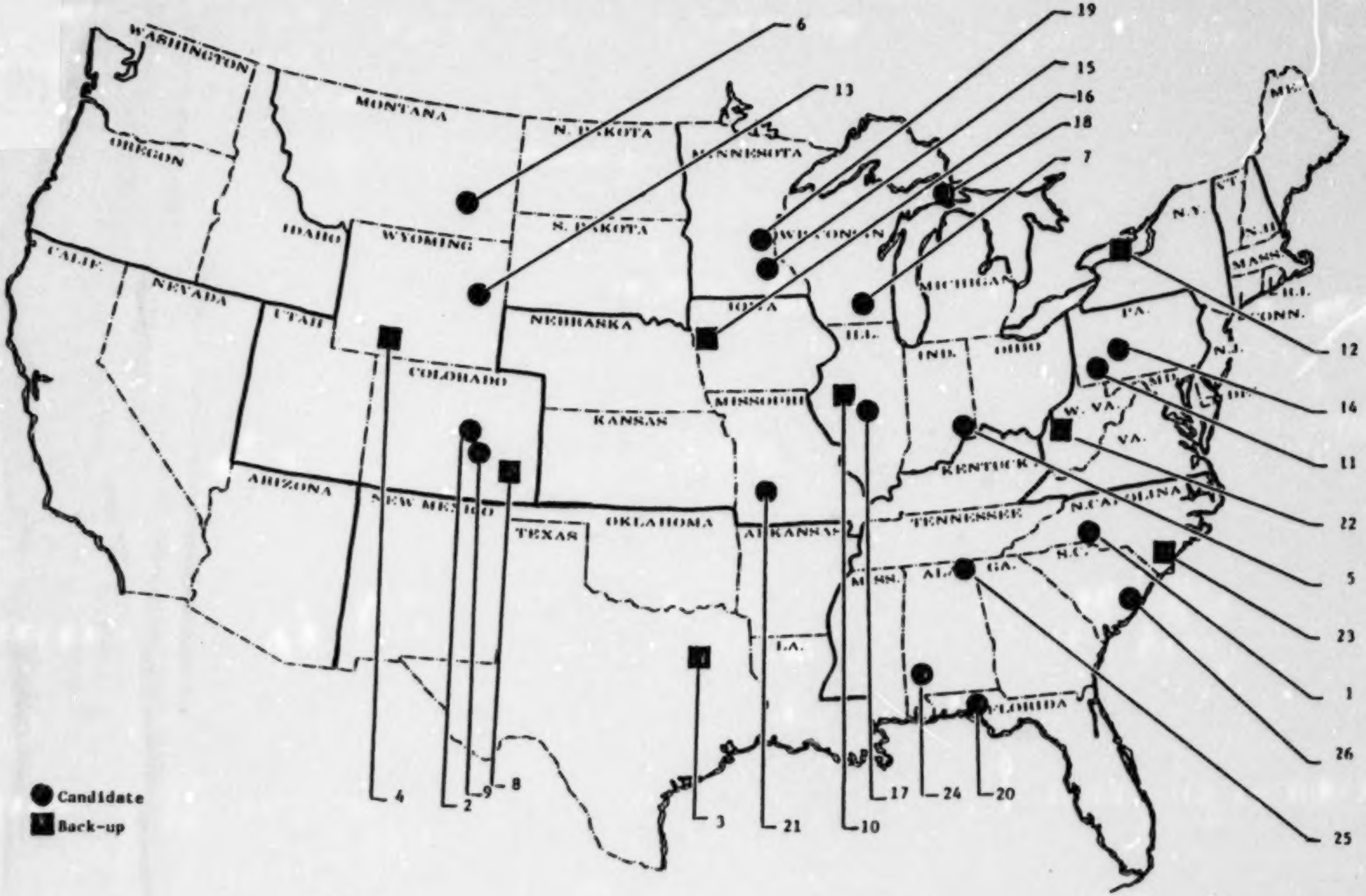
- Major variations in geotechnical and environmental factors within a region might require the selection of more candidate sites for that region. Similarly, the converse might be true in other regions.

- Projections of future coal utilization and trends in disposal in the United States strongly emphasize the striking growth in production and utilization of western coal. Hence it would be appropriate to consider more candidate sites from the western part of the country and more plants elsewhere that may use western coal in the future. With revised NSPS regulations for utility boilers requiring scrubbing and/or other sulfur control for all coal (including low-sulfur coal), the potential attractiveness of western coal as a way to avoid sulfur control in eastern regions has changed. However, western coal is likely to be utilized at an ever increasing rate.

3.2.3 Initial Screening of Sites

The initial screening process consisted of:

- Definition of overall study objectives. The overall objective was to select sites that would include the prevalent methods of disposal used by industry today and those with future potential. For example, it appears that in the future dry disposal of FGC wastes (that is, as moist soil-like materials) in managed fills


WASHINGTON
OREGON
CALIF.
NEVADA
IDAHO
MONTANA
WYOMING
UTAH
COLORADO
ARIZONA
NEW MEXICO
N. DAKOTA
S. DAKOTA
NEBRASKA
KANSAS
OKLAHOMA
TEXAS
MINNESOTA
IOWA
MISSOURI
ARKANSAS
LA.
WISCONSIN
ILL.
MICHIGAN
IND.
OHIO
KENTUCKY
TENNESSEE
MISS.
ALA.
GA.
FLORIDA
S.C.
N.CAROLINA
VA.
W. VA.
PA.
N.Y.
N.J.
MD.
DE.
VT.
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1
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9
8
3
21
10
17
24
20
● Candidate
■ Back-up

Note: See Table 3.1 for Map Key

FIGURE 3.1 LOCATION OF CANDIDATE AND BACKUP SITES

TABLE 3.1

MAP KEY FOR LOCATION OF CANDIDATE AND BACKUP SITES (FIG. 3.1)

Map Number	Plant Name	Utility Name	County	State
1	Allen	Duke Power	Gaston	NC
2	Arapahoe	Public Service of Colorado	Denver	CO
3	Big Brown (Backup)	Texas Power & Light	Freestone	TX
4	Jim Bridger (Backup)	Pacific Power & Light	Sweetwater	WY
5	Clifty Creek	Indiana-Kentucky Electric	Jefferson	IN
6	Colstrip	Montana Power	Rosebud	MT
7	Columbia I	Wisconsin Power & Light	Portage	WI
8	Comanche (Backup)	Public Service of Colorado	Pueblo	CO
9	Drake	Colorado Springs Dept. Pub. U.	El Paso	CO
10	Duck Creek (Backup)	Central Illinois Light	Fulton	IL
11	Elrama	Duquesne Light	Washington	PA
12	Huntley (Backup)	Niagara Mohawk Power	Erie	NY
13	Dave Johnston	Pacific Power & Light	Converse	WY
14	Keystone	Pennsylvania Electric	Armstrong	PA
15	A.S. King	Northern States Power	Washington	MN
16	George Neal (Backup)	Iowa Public Service	Woodbury	IA
17	Powerton	Commonwealth Edison	Tazewell	IL
18	Presque Isle	Upper Peninsula Generating	Marquette	MI
19	Sherburne County	Northern States Power	Sherburne	MN
20	Smith	Gulf Power	Bay	FL
21	Southwest	Springfield City Utilities	Greene	MO
22	Sporn (Backup)	Central Operating	Mason	WV
23	Sutton (Backup)	Carolina Power & Light	New Hanover	NC
24	Tombigbee	Alabama Electric Coop.	Washington	AL
25	Widows Creek	Tennessee Valley Auth.	Jackson	AL
26	Winyah	S. Carolina Public Ser. Auth.	Georgetown	SC

is likely to be encouraged. Similarly, use of coal in western plants is likely to grow. Factors such as the above were recognizable in defining required characteristics of suitable sites. In defining study objectives it is well to note that the focus of this study is those subject areas that are priority concerns under RCRA for utility solid waste disposal. These are impacts on groundwater quality, impacts on surface water quality from non-point sources, and use of mitigative design, control, or management practices. The subsequent criteria for selection were based on the need to study impacts in these areas.

- Development of appropriate engineering/technology-related screening criteria to assess a plant's disposal operations. Engineering/technology criteria included factors related to waste variability, site age, FGD system, and application of mitigative engineering practices. These were applied as the first-level screens.

- Development of hydrogeologic criteria. The hydrogeologic constraints related to bedrock geology, surficial soil characteristics, groundwater flow conditions at the site, and climate. The goal was to select sites where data from about 1 year environmental monitoring could be used for assessment with reasonable confidence.

- Development of other site selection criteria, taking into account overall objectives of RCRA.

The above screening criteria (engineering/technological, hydrogeologic, and other site selection factors) were used to obtain a preliminary list of sites where reasonable assessment of data obtained from about 1 year of monitoring would be possible. The approach was to focus on sites where cost-effective data gathering and reliable assessment of data were possible. The overall objective of this project is to provide the information necessary to develop reasonable guidelines for all kinds of disposal sites, including very complex ones. However, for the research phase of this effort, it is necessary to select sites where data from 1 year of monitoring can be used for reliable assessment. Therefore, very complex sites where reliable data interpretation is not possible (on technological or hydrogeologic grounds) were eliminated from further consideration. However, the overall plan for characterization and environmental monitoring at the final 12 sites will be designed to permit assessment and development of recommendations applicable to complex sites, complex waste types, and complex methods of disposal.

3.2.4 Final Ranking of Sites

The investigators evaluated the initial list of sites in an iterative manner taking into account three categories of variables:

- Waste Type: An analysis of the available information indicated that the disposal of six categories of wastes constitutes the vast majority--fly ash disposal, bottom ash disposal, combined fly and bottom ash disposal, disposal of FGD waste alone, FGD waste and fly ash codisposal, and stabilized FGD waste disposal. These combinations are likely to remain the only major options of importance in the future.

- Method of Disposal: Ponding and landfill type operations are practicable for many wastes. However, stabilized disposal of FGD waste as moist soil-like material in landfills is likely to be very important. Similarly, FGD wastes, when disposed of alone, are usually disposed of by ponding only. In the future, with forced oxidation methods, FGD waste disposal by itself in the form of gypsum stacks, may be practical; however, such disposal practices do not exist on full scale at present. Hence, FGD waste disposal by ponding alone was considered. Similarly, mine disposal is a special subcategory of landfill type disposal and was considered when appropriate.

- General Location: While the initial screening was done on the basis of 14 discussion regions, the final ranking for presentation use was based on dividing the United States into 3 environmental zones: Coastal, Interior, and West. Plants within a few miles of Atlantic and Gulf Coasts were grouped as Coastal. Plants in or west of Montana, Wyoming, Colorado, and New Mexico were considered Western. The remainder of the contiguous 48 states were considered Interior.

The final ranking process began by listing all sites with a particular mix of waste type, method of disposal (e.g., fly ash ponding), and general location. The sites were ranked relative to each other and the more promising ones were recommended as candidate sites, the others as backups. In some cases backups were not available and in other cases backups were in different regions from the candidate. Collectively, the backup sites provide the capability to measure effects if some of the candidates are found unsuitable. The final results are shown in:

- Figure 3.1 showing location of sites,

- Table 3.1 showing names and locations,

- Table 3.2 showing data on the plants, and

- Table 3.3 showing the matrix of waste type, disposal method, and general location.

The recommended list of candidate and backup sites provides a broad mix of options to be subjected to further analysis prior to the selection of the final 12 sites. Detailed site visits, together with an evaluation

TABLE 3.2

RECOMMENDED CANDIDATE AND BACKUP SITES

	Candidate			Backup		
	Coast	Interior	West	Coast	Interior	West
1 Fly Ash: Pond	NA	Columbia I Clifty Creek	AT	NA	- Sporn	AT
Landfill	AT	Powerton (AL) Presque Isle	D. Johnston	AT	Big Brown Huntley	Comanche
2 Fly/Bottom Ash: Pond	Smith	Elrama (I) Widows Creek Allen	Arapahoe (I) D. Johnston (I)	Sutton	G. Neal	-
Landfill	AT	Keystone A. S. King	Drake	AT	-	J. Bridger (M)
3 Fly Ash/FGD Waste: Pond	NA	Sherburne County	Colstrip (DP)	NA	Duck Creek	-
Landfill	AT	Southwest (AL)	AT	AT	-	AT
4 Stabilized FGD Waste:	NA	Elrama	NA	NA	-	NA
Landfill	NA	Elrama	NA	NA	-	NA
5 FGD Waste: Pond	Winyah Tombigbee (L)	Widows Creek	AT	-	-	AT
6 Bottom Ash: Pond	AT	Sherburne County Columbia I Clifty Creek Presque Isle (I) Southwest (I) (AL) Keystone (I)	Colstrip (DP) D. Johnston (I)	AT	Big Brown Duck Creek Huntley (I)(AL) Sporn	Comanche (I)
Landfill	AT	Powerton (AL)	AT	AT	-	AT

(AL) = Artificially Lined

(DP) = Two ponds in series

(AT) = Atypical disposal practice for region

(NA) = None available at present

(I) = Interim Pond/Landfill (interim/final)

(L) = Lined (exact liner unknown)

(M) = Mine

TABLE 3.3

RECOMMENDED CANDIDATE AND BACK UP SITES - OVERVIEW

No.	Plant Name	Utility Name	Location		Nameplate Generating Capacity (MW)	Start-up Date[a]	Disposal	Region
			County	State				
	CANDIDATE SITES							
1	Allen	Duke Power	Gaston	NC	1155	-/57	Combined fly and bottom ash to an unlined pond.	Interior
2	Arapahoe	Public Service of Colorado	Denver	CO	250	-/50	Combined fly and bottom ash to an interim pond and then to a landfill.	West
3	Clifty Creek	Indiana-Kentucky Electric	Jefferson	IN	1303	2/55	Fly and bottom ash to separate clay substrate lined ponds.	Interior
4	Colstrip	Montana Power	Rosebud	MT	720 (720 on FGD)	11/75	Fly ash/FGD wastes to a clay substrate lined interim pond and then to an unlined final pond. Bottom ash to separate clay substrate lined pond interim pond and then to the same final pond.	West
5	Columbia I	Wisconsin Power & Light	Portage	WI	556	5/75	Fly and bottom ash to separate unlined ponds.	Interior
6	Drake	Colorado Springs Dept. of Public Utilities	El Paso	CO	282	-/62	Combined fly and bottom ash to a clay substrate lined landfill.	West
7	Elrama	Duquesne Light & IU Conversion Systems	Washington	PA	510 (510 on FGD)	6/52 FGD in 10/75	FGD wastes stabilized and disposed of in an off-site landfill. Combined fly and bottom ash to an unlined interim pond and then to the same landfill.	Interior
8	Dave Johnston	Pacific Power & Light	Converse	WY	750	-/59	Fly ash to an unlined landfill. Combined fly and bottom ash to a clay substrate lined interim pond and then to the same landfill. Bottom ash to an unlined interim pond and then to the same landfill.	West
9	Keystone	Pennsylvania Electric	Armstrong	PA	1872	1/67	Fly ash to an unlined landfill. Bottom ash to an unlined interim pond and then to the same landfill.	Interior
10	A. S. King	Northern States Power	Washington	MN	598	-/68	Combined fly and bottom ash to an unlined landfill.	Interior

TABLE 3.3 (Continued)

No.	Plant Name	Utility Name	Location		Nameplate Generating Capacity (MW)	Start-up Date[a]	Disposal	Region
			County	State				
11	Powerton	Commonwealth Edison	Tazewell	IL	1785	-/72	Fly ash to an artificially lined landfill. Bottom ash to the same landfill. Landfill is off-site.	Interior
12	Presque Isle	Upper Peninsula Generating	Marquette	MI	301	9/55	Fly ash to an unlined landfill. Bottom ash to an unlined interim pond and then to the same landfill.	Interior
13	Sherburne County	Northern States Power	Sherburne	MN	1440 (1440 on FGD)	5/76	Fly ash/FGD wastes to a clay substrate lined pond. Bottom ash to an interim pond and then sold.	Interior
14	Smith	Gulf Power	Bay	FL	340	6/65	Combined fly and bottom ash to an unlined pond.	Coastal
15	Southwest	Springfield City	Greene	MO	194 (194 on FGD)	6/76 FGD in 4/77	Fly ash/FGD wastes to an artificially lined landfill. Bottom ash to an artifically lined interim pond and then to the same landfill.	Interior
16	Tombigbee	Alabama Electric Coop.	Washington	AL	585 (385 on FGD)	6/69 FGD in 9/78	FGD wastes to a lined pond.	
17	Widows Creek	Tennessee Valley Authority	Jackson	AL	1977 (550 on FGD)	7/52 FGD in 1/78	FGD wastes to an unlined pond. Combined fly and bottom ash to an unlined pond.	Interior
18	Winyah	South Carolina Public Service Authority	Georgetown	SC	630 (140 on FGD)	5/75 FGD in 7/77	FGD wastes to an unlined pond. Combined fly and bottom ash probably ponded.	Coastal
	BACKUP SITES							
19	Big Brown	Texas Power & Light	Freestone	TX	1186	12/71	Fly ash to a clay substrate lined landfill. Bottom ash to clay substrate lined pond.	Interior
20	Jim Bridger	Pacific Power & Light	Sweetwater	WY	1525	9/74	Combined fly and bottom ash to a landfill which is a mine.	West

TABLE 3.3 (Continued)

No.	Plant Name	Utility Name	Location		Nameplate Generating Capacity (MW)	Start-up Date[a]	Disposal	Region
			County	State				
21	Comanche	Public Service of Colorado	Pueblo	CO	778	-/73	Fly ash to a landfill. Bottom ash to an interim pond and then to the same landfill.	West
22	Duck Creek	Central Illinois Light	Fulton	IL	441 (416 on FGD)	6/76 FGD in 9/76	Fly ash/FGD wastes to an unlined pond. Bottom ash to an unlined pond.	Interior
23	Huntley	Niagara Mohawk Power	Erie	NY	828	-/42	Fly ash to an unlined landfill. Bottom ash to unlined and lined interim ponds and then to the same landfill.	Interior
24	George Neal	Iowa Public Service	Woodbury	IA	961	-/64	Combined fly and bottom ash to an unlined pond.	Interior
25	Sporn	Central Operating	Mason	WV	1105	-/50	Fly ash to an unlined pond. Bottom ash to an unlined interim pond and then sold.	Interior
26	Sutton	Carolina Power & Light	New Hanover	NC	671	8/54	Combined fly and bottom ash to an unlined pond.	Coastal

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[a] Original start-up date of plant's first coal-fired unit.

of the available environmental and hydrogeologic information, are required to carry this process further. Additionally, in each discussion region, other potentially attractive sites have been identified, thus providing a second level of backup beyond the 26 recommended sites. With this overall understanding, the candidate and backup sites described in the Site Selection Report were recommended for further consideration. Detailed evaluations of the candidate and backup sites are in progress.

Since June 1980, the candidate and backup sites have been subjected to further evaluation to arrive at the final 12 sites. Evaluation efforts have included:

- Discussions with utility personnel to check the validity of our data base and field visits by the team of investigators to verify and enlarge our data base.
- Where appropriate, full hydrogeologic and geotechnical evaluations of the sites to provide first-hand and detailed accounts of the site hydrogeology.
- Upon approval of EPA, preparation of preliminary test plans on the selected sites.

As of August 1980, Plant Allen of Duke Power (Charlotte, NC) has been recommended by ADL to be a selected site. By late October 1980, at least three other sites are likely to be recommended.

3.3 Planning of Procedures Manuals

Four procedures manuals have been planned to precisely define the procedures to be employed in field investigations: site development, sampling and analysis, engineering/economic assessment, and environmental assessment. These manuals will serve as baseline documents for preparation of test plans at the 12 final sites and the conduct of the characterization and environmental monitoring program. They will also be included as appendices in each of the 12 test plans. The first two manuals have been completed in draft form and all of the draft manuals will be completed by November 1980. A brief review of these is presented below.

- Appendix A - Hydrogeologic and Geotechnical Procedures Manual

  This manual has been prepared in draft and will serve as the basic compendium for the conduct of hydrogeologic and geotechnical investigations on the sites (the procedures and methods to be employed for the development of field wells and other aspects of hydrogeologic and geotechnical considerations). It should be of great value to anyone who considers a monitoring program, be it a utility, EPRI, or those involved in other kinds of field monitoring of waste disposal operations.

- Appendix B - Sampling and Analysis Procedures Manual

  This manual is also completed in draft and covers the following subjects:

  - Sampling of wastes, soil, groundwater, and surface water.
  - Physical and engineering characterization of appropriate materials.
  - Chemical characterization of the above materials.

- Appendix C - Engineering/Economic Assessment Manual

  This manual is under development and will include baseline approaches for engineering and economic assessment. It is expected that the draft manual will be completed by November 1980. The overall approach has been to define uniform procedures to obtain capital and operating costs on the same basis for different disposal operations constructed at different times in different parts of the country. Additionally, the use of information on reliability and reproducibility of disposal operations in determining costs will be included in the manual.

- Appendix D - Environmental Assessment Manual

  This manual is in preparation and its draft will be available in early December. The environmental analysis approach will be clearly developed in this manual.

3.4 Utility Cooperation

EPA and ADL have been successful to date in fostering good communication, through the mechanism of the two committees, with the utility associations and a number of utilities. Their close participation and review of major documentation has been a substantial factor in ensuring cooperation. A high level of cooperation has been noted and it is very likely due to the interactions that have taken place.

3.5 Anticipated Accomplishments in the Next 18 Months

It is estimated that the project will be completed and the final report submitted by November 1982. The accomplishments anticipated over the next 18 months (i.e., until the next FGD Symposium, hopefully in early 1982) are:

- Over the next 8 to 10 months all 12 sites will have been selected and developed, and monitoring will have begun.
- By later 1981, initial baseline data will be adequate for an interim report to EPA.
- Any unusual problems will have been highlighted and resolved.

4.0 REGULATORY/LEGISLATIVE DEVELOPMENTS

4.1 RCRA Regulatory Developments

RCRA requires that EPA establish such standards as are necessary to protect human health and the environment. The field studies to be performed under this project are designed to provide sufficient data to evaluate the performance of various disposal activities, practices, etc., with respect to their effects and potential for impact on human health and the environment. The regulations for hazardous waste under RCRA which were issued in May 1980 and recently passed Congressional amendments provide a temporary deferral for fly ash, bottom ash, boiler slag, and flue gas emission control waste and exclude them from any RCRA hazardous waste regulations pending completion of this project and other studies. It is EPA's intent that this project partially fulfill requirements of these studies.

The two major factors that need to be considered in any prognosis on the hazardous/nonhazardous questions are:

- RCRA toxicity criteria for waste extracts from Section 3001 (EP) have now been changed from 10 times the relevant drinking water standards to 100 times those standards.
- Based on the best information available to date, EP extracts from coal ash and FGD wastes of the utility industry are expected to be well below 100 times the standards.

On this basis, it appears that utility solid wastes would fall in the nonhazardous category and that EPA will ultimately write guidelines for Federal and state permitting officials concerned with regulation of the disposal of utility solid wastes. Nevertheless, EPA has directed ADL to gather grab samples of wastes from every site evaluated to obtain data on levels of specific contaminants in the extracts using the RCRA Section 3001 protocol. This process is underway. It is anticipated that this work will confirm the preliminary expectation that utility waste extracts will be well below the toxicity criteria. However, ADL has been directed to schedule the analyses so that, if data collected under the project warrants, the information developed using the Section 3001 protocol could be used as one of the criteria for site selection.

4.2 RCRA Legislative Developments

Two bills have been passed by the House and Senate mandating that EPA undertake studies concerning specific aspects of coal ash and flue gas desulfurization wastes. These are HR 3994, passed on February 20, 1980, and S 1156, passed on June 4, 1979. A compromise bill was reported out of conference committee in early October and is expected to become law in late 1980. It is anticipated that the enacted version will require that EPA undertake a detailed and comprehensive study on the adverse effects, if any, on human health and the environment, of the disposal or utilization methods employed for fly ash, bottom ash, flue gas emission control waste, and other by-product materials generated primarily from the combustion of coal or other fossil fuel. It is expected that the Congressional mandate will require an analysis of:

- The source and volume of such materials generated per year.
- Present disposal and utilization practices.
- Potential danger, if any, to human health and the environment from the disposal and reuse of such materials.
- Documented cases in which danger to human health or the environment from surface runoff or leachate has been proven.
- Alternatives to current disposal methods.
- The cost of such alternatives.
- The impact of those alternatives on the use of coal and other natural resources.
- The current and potential utilization of such materials.

Many of the items required under the above eight categories will be addressed by this project. Additional effort is also anticipated by EPA itself in providing the comprehensive information required by this bill.

4.3 Developments Under Power Plant and Industrial Fuel Act of 1978

Under this law, the Department of Energy has been evaluating potential changeover from other fossil fuels to coal in various parts of the country. To date, 13 individual orders proposing such potential switches to coal have been issued by DOE. This is likely to continue in the future. It is estimated by the Department of Energy that ultimately up to 115 plants are likely to be under such consideration (5).

Such changeovers are particularly likely in areas of the country such as New England where coal-fired plants have been relatively few in the past. The utility industry has been interested in knowing whether a site of such changeover to coal firing may be selected under this project. Implementation of such changeovers is likely to be at least 1 or 2 years or more in the future. Hence, such sites cannot be included in this project. However, steps are being taken to ensure that a representative mix of waste types, disposal practices, and site-specific hydrogeologic and climatic characteristics will be studied at selected sites in other parts of the country so that projections of potential impacts of coal-fired power plants in New England and other regions of the country where conversion to coal is likely will be possible. (Note that New England is one region where at-sea disposal of FGC wastes, if practiced in an environmentally sound manner, may be attractive.)

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EVALUATION OF POTENTIAL IMPACTS TO THE UTILITY SECTOR FOR COMPLIANCE WITH RCRA

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ABSTRACT:

This paper presents "interim" findings of a continuing evaluation of the impacts on coal-fired electric generating facilities in the utility sector deriving from Proposed Rules issued by EPA for implementing Sections 3001, 3002, and 3004 of RCRA Subtitle C (18 December 1978) and Proposed Guidelines under Section 4004 (6 February 1979) and Section 1008 (26 March 1979). Cost analyses were made on direct costs only and presented in 1979 dollars.

The primary purpose for undertaking this study was to assess RCRA's impacts on coal-fired utilities as a means to understanding the implications of RCRA regulations on the National Energy Plan (NEP), the Powerplant and Industrial Fuel Use Act of 1978 (PIFUA), which gives DOE authority to enforce coal conversion actions, and on utilities. Ultimately, this information can be used to understand and determine RCRA's implications on the Department of Energy, Fossil Energy, mission objectives of developing programs and supporting advanced technologies which will promote greater utilization of coal in an environmentally acceptable manner.

Cost and other impact considerations were developed through conduct of a survey of available information from a total of 29 significant coal producing and/or using states and by conducting in-depth case studies of operating conditions and engineering and design requirements for RCRA compliance at six operating and 16 theoretical base-case coal-fired power generating plants. The six plants were selected to encompass a broad cross-section of industry operating factors. Solid waste disposal costs were developed for three scenarios structured to bracket the probable cost impacts of the proposed RCRA Regulations on the utility sector. Cost curves and mathematical models of capital and operation and maintenance (O&M) costs associated with each disposal scenario were developed as a function of waste generation rates. In all instances the "least cost of compliance" rather than "worse case" data were employed.

Preliminary national costs for waste disposal, based on direct costs alone, reflected an increase in disposal costs of over "4 times" for non-hazardous and over "13 times" for hazardous disposal. This results in a cost increase of from 1.09 to 3.2 mils per kwh of power generated (in terms of 1979 dollars).

EVALUATION OF POTENTIAL IMPACTS TO THE UTILITY SECTOR FOR COMPLIANCE WITH RCRA

In an effort toward achieving the strategic national objective of energy independence, the National Energy Plan (NEP) has placed strong emphasis on greater development and increased utilization of coal. The Powerplant and Industrial Fuel Use Act of 1978 (PIFUA) has been the recent authority under which the Department of Energy (DOE) has sought to implement the goals of the NEP through enforcement of conversion of utility and industrial boilers from oil and natural gas to coal and coal-based fuels. The increased coal use mandated by the PIFUA will impose greater requirements for the proper disposal or management of high volume coal combustion solid wastes and residues than for oil or natural gas waste products.

To achieve an understanding of the implications of greater coal utilization and the concomitant solid waste problems, the Office of Coal Utilization, DOE, retained Engineering-Science to evaluate the impact on coal-fired facilities resulting from proposed regulations issued by the U. S. Environmental Protection Agency (EPA) for implementation of the Resource Conservation and Recovery Act of 1976 (RCRA).

Most of the information presented in this paper is based upon preliminary findings as published in the DOE Interim Report, Phase I - Utility Sector, published in November 1979 and entitled, Evaluation of the Impacts of the Proposed RCRA Regulations on Utility and Industrial Sector Fossil Fuel-Fired Facilities. The hazardous disposal cost data assessment of the utility sector is based upon Proposed Rules issued by EPA for implementing Sections 3001, 3002, and 3004 of RCRA Subtitle C (18 December 1978), and the nonhazardous cost data were based on Proposed Guidelines under Section 4004 (6 February 1979), and Section 1008 (26 March 1979). Publication recently of parts of of the final RCRA regulations notwithstanding, the author believes much of the data presented here is still germane for purposes of evaluation since EPA has not yet published its final versions of the Section 3001 and 3004 standards. Further, there is still uncertainty as to what ultimate level of utility waste regulation will be recommended by EPA several years from now at the conclusion of its research on utility waste disposal at a number of study sites. Other key assumptions used during the development of data can be summarized as follows:

- Cost analyses were made on direct costs only, presented in 1979 dollars.

- Waste disposal sites were located so as to minimize construction and operating costs.

- Regional and national disposal cost estimates were based on evaluation of one coal-fired generating station in each of six selected Federal Energy Regulatory Commission (FERC) regions. Waste disposal costs for each powerplant were assumed to be representative of costs of all plants in that region. (The range and variety of cost factors for any given region will be explored in the final report.)

- A weighted average of disposal costs for the six case study plants was assumed to be representative of the nationwide costs.

- Wastes considered in the study were fly ash, bottom ash and FGD sludge with no provisions for segregation or separate disposal.

- Only coal-fired electric generating facilities with greater than 25 MW capacity were considered in the study.

- State and local jurisdictional impediments, such as could occur with transportation of wastes across county and state lines, were not included in the cost evaluations.

These assumptions were necessitated by the guidance of DOE to the contractor that he conduct the study in such a way as to make conservative estimates of the impacts of the proposed RCRA regulations, i.e., the lowest cost of compliance.

The primary purpose for undertaking the study was to assess RCRA's impacts on coal-fired utilities as means to understanding the implications of RCRA regulations on the National Energy Plan, the Powerplant and Industrial Fuel Use Act of 1978, which gives DOE authority to enforce coal conversion actions, and on utilities. Ultimately, this information can be used to understand and determine RCRA's implications on the Department of Energy, Fossil Energy mission objectives of developing programs, and supporting advanced technologies which can promote greater utilization of coal in an environmentally acceptable manner. Specifically, the project originally had four major objectives:

1. Analyze RCRA impacts on coal-fired utilities.

2. Assess implications on emerging coal technologies; examine the comparative economic viability of various emerging coal technologies in light of potential waste management compliance requirements under RCRA.

3. Assess implications on implementation of the Powerplant and Industrial Fuel Use Act of 1978.

4. Develop a range of potential national impacts which might be experienced by implementation of RCRA on major coal-fired facilities.

Today, the objectives of the study go well beyond RCRA and are aimed at examining the range of impacts that derive from RCRA and its interactions with a host of other environmental regulations, which will have an interfacing influence on RCRA.

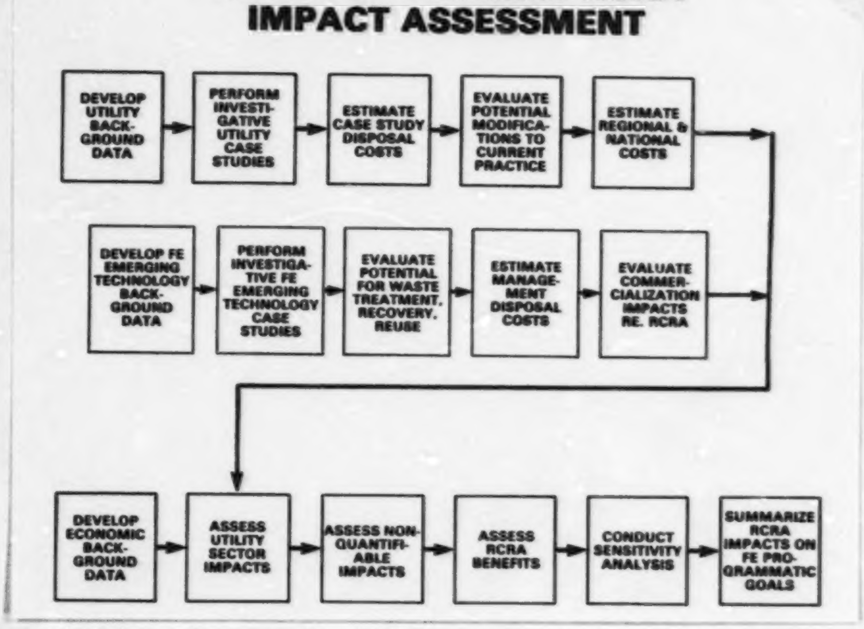
The project is being conducted as a three phase effort. Phase I deals with the utility sector; Phase II deals with the emerging coal technologies (advanced combustion, coal conversion and advanced environmental control); Phase III consists of an analysis of financial, economic, institutional and programmatic impacts on a nationwide basis. This paper presents an overview of the methodology used for the entire project as well as a summary of results obtained to date.

Phase I

Work completed to date consists primarily of an analysis of potential costs to the utility sector based on the RCRA regulations proposed by EPA on December 18, 1978. Work is presently in progress to update these cost estimates in accordance with the final RCRA regulations promulgated in 1980 and to increase the accuracy of these estimates by expanding the data base.

The scope of activities undertaken to accomplish the project objectives is shown in Figure 1, along with the sequence and interrelationships among the various major tasks.

Figure 1


SCOPE OF CURRENT RCRA IMPACT ASSESSMENT
DEVELOP UTILITY BACKGROUND DATA
PERFORM INVESTIGATIVE UTILITY CASE STUDIES
ESTIMATE CASE STUDY DISPOSAL COSTS
EVALUATE POTENTIAL MODIFICATIONS TO CURRENT PRACTICE
ESTIMATE REGIONAL & NATIONAL COSTS
DEVELOP FE EMERGING TECHNOLOGY BACKGROUND DATA
PERFORM INVESTIGATIVE FE EMERGING TECHNOLOGY CASE STUDIES
EVALUATE POTENTIAL FOR WASTE TREATMENT, RECOVERY, REUSE
ESTIMATE MANAGEMENT DISPOSAL COSTS
EVALUATE COMMERCIALIZATION IMPACTS RE. RCRA
DEVELOP ECONOMIC BACKGROUND DATA
ASSESS UTILITY SECTOR IMPACTS
ASSESS NON-QUANTIFIABLE IMPACTS
ASSESS RCRA BENEFITS
CONDUCT SENSITIVITY ANALYSIS
SUMMARIZE RCRA IMPACTS ON FE PROGRAMMATIC GOALS

The contractor, in pursuit of the various task responsibilities, initiated compilation of a massive data base, reviewed existing literature and Federal reporting activities, and surveyed, independently and in conjunction with the National Governors' Association, 29 significant coal producing and using states. Information was collected regarding waste generation quantities, current waste disposal methods employed by utilities, existing state solid waste disposal regulations, and projections of the near-term increases in coal burning and waste generation.

More specific information was obtained by selecting six operating coal-fired power generating plants for in-depth case studies. These case study plants were selected in order to obtain a broad cross-section of the industry with respect to such factors as geographic location, generating capacity, solid emission controls as well as other significant factors. The following plants were thus selected by DOE in cooperation with representatives from the utility sector as case study sites:

Plant Name	FERC Region	Generating Capacity	Type of Coal
Bowen (GA)	5	3160 MW	Bituminous
Conesville (OH)	3	1981 MW	Bituminous
Tombigee (AL)	6	585 MW	Bituminous
Eddystone (PA)	2	1489 MW	Bituminous
Martin Lake (TX)	7	2379 MW	Lignite
Colstrip (MT)	8	716 MW	Sub-Bituminous

One or more site visits were made to each plant to verify detailed information regarding the type of power generating equipment, source of coal, type of air emission controls, and current solid waste disposal practices. Additional site-specific information regarding physical characteristics such as topography, subsurface geology, groundwater elevations, and non-quantifiable impacts such as jurisdictional problems, local socio-economic conditions, and public attitude were also obtained from various sources. This and other information served as the basis for developing engineering cost models from which estimates of regional and national cost impacts could be evolved relative to proposed RCRA regulations. It should be noted that the six case study sites do not necessarily represent a statistically valid sampling for estimation of regional or national cost impacts. However, this first phase of impact analyses was developed to obtain only a preliminary assessment of the associated impacts of proposed regulations under RCRA. After the project has progressed sufficiently, a sensitivity analysis will be conducted on the various elements making up the overall assessment to determine the range of implications of RCRA and other related Federal/environmental regulations on the mission objectives of the Fossil Energy (FE) Program components.

Integrated into the above activities is consideration of a separate but simultaneous parallel evaluation and assessment of RCRA and other related environmental regulatory implications relative to FE emerging technolgoy projects. One of the key motivations here is to examine the impact offset benefits and potential issues associated with waste stabilization, fixation, resource recovery, and reutilization applications of coal combustion (and/or conversion) waste products and residues.

Before explaining the actual cost and impact estimating methodology, it is interesting to note some of the DOE observations made on the National Governors' Association State Survey, and on data acquired from State Offices of Solid Waste. This information is summarized below in Figures 2 and 3.

Figure 2

OBSERVATIONS OF THE NATIONAL GOVERNORS' ASSOCIATION STATE SURVEY*:

- ONLY NEW YORK, PENNSYLVANIA AND WEST VIRGINIA HAVE DISPOSAL REGULATIONS SPECIFICALLY FOR COAL-FIRED UTILITY WASTES.
- PROPOSED RCRA SECTION 3004 REGULATIONS ARE MUCH MORE STRINGENT THAN MOST EXISTING STATE REGULATIONS.
- THE STATES REPORTED THAT THE MAJORITY OF EXISITNG SOLID WASTE DISPOSAL SITES GENERALLY CONFORM TO SECTION 4004 GUIDELINES AND MOST OF THE REMAINING SITES MAY BE UPGRADED TO MEET SECTION 4004 GUIDELINES. HOWEVER, ENGINEERING-SCIENCE'S INDEPENDENT EVALUATIONS DO NOT SUPPORT THIS OBSERVATION.

*AS REPORTED IN THE E-S INTERIM REPORT (NOVEMBER 1979).

Figure 3

INFORMATION NOT READILY AVAILABLE FROM THE STATE OFFICES OF SOLID WASTE

- PERCENTAGE OF ON-SITE VERSUS OFF-SITE DISPOSAL
- METHOD OF OFF-SITE TRANSPORTATION AND PERCENTAGE OF TOTAL OFF-SITE TRANSPORTATION
- OFF-SITE TRANSPORTATION COSTS
- OFF-SITE HAUL DISTANCES
- AVAILABLE CAPACITY OF EXISTING SOLID WASTE DISPOSAL SITES
- SPECIFIC REQUIREMENTS NEEDED TO UPGRADE CURRENT DISPOSAL SITES TO RCRA 3004 AND 4004 CRITERIA
- COSTS OF COMPLIANCE FOR UPGRADING EXISITNG DISPOSAL SITES TO MEET RCRA 3004 AND 4004 CRITERIA
- PERCENTAGE OF LAND AREA LOCATED IN ENVIRONMENTALLY SENSITIVE AREAS OR OTHER AREAS UNSUITABLE FOR DISPOAL SITES

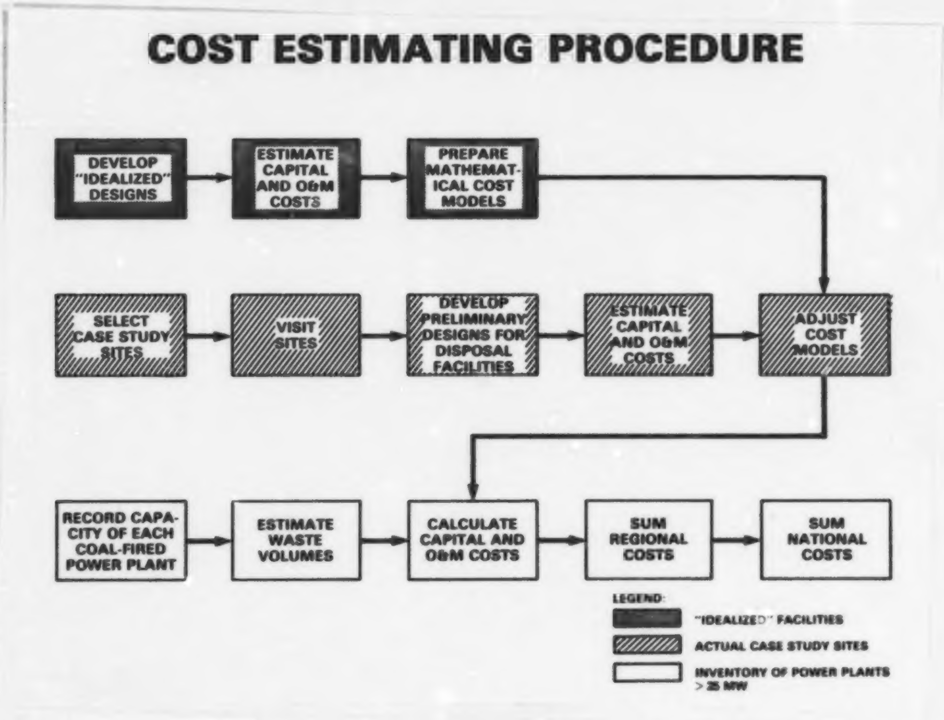
The reporting of the states on the conformance of most of their solid waste disposal sites to RCRA, Section 4004 guidelines have proven to be generally erroneous. Many of these sites are nothing more than excavations or unimproved landfills lacking liners or provision for collection of leachate. For the most part, until recently, most states had demonstrated that they were ill-informed of the regulations and ill-prepared to assess their existing solid waste situation or to evaluate their future requirements in an orderly and consistent fashion.

The lack of credible basic informaiton, as illustrated above from most of the states, can burden the progress of a study such as this one, and has necessitated going to the "grass roots level," whenever possible, to try to develop accurate original data. In some instances, Federal data may be no better than that at the state level as demonstrated by the fact that the author had the contractor examine data from the Federal Power Commission Form 67, which is required under Federal law of each power generating facility in the country. Considering 65 plants in the case study states alone, only 6 reported even partial or incomplete data on these forms!

As indicated above, Phase I of the project consists of an assessment of potential RCRA impacts on the utility sector. The overall approach has been based on having the contractor, Engineering-Science (E-S) assume the role of a consultant retained by various utilities for the purpose of designing waste disposal facillities and estimating their costs. The author has been committed to the position that in developing credible cost/impact data, DOE must depart in this project from the typical and traditional "paper study" format that is so popular in Government and that is only as credible as the weakest links in the long chain of other paper studies used as the foundation upon which new premises are built. Therefore, the cost estimating procedure we implemented emulates the typical approach used by industry. Extensive engineering evaluations at the various sites were conducted, hard preliminary engineering designs for facilities to meet specific RCRA compliance requirements were developed, and those requirements were costed out based on actual cost experience in 1979 dollars the same way industry would proceed with development of a facility. This "hands on" approach was selected because it was felt that the results would best reflect the true costs of implementing the regulations.

Figure 4 below presents the various steps utilized in the cost estimating process.

Figure 4


COST ESTIMATING PROCEDURE
DEVELOP "IDEALIZED" DESIGNS
ESTIMATE CAPITAL AND O&M COSTS
PREPARE MATHEMATICAL COST MODELS
SELECT CASE STUDY SITES
VISIT SITES
DEVELOP PRELIMINARY DESIGNS FOR DISPOSAL FACILITIES
ESTIMATE CAPITAL AND O&M COSTS
ADJUST COST MODELS
RECORD CAPACITY OF EACH COAL-FIRED POWER PLANT
ESTIMATE WASTE VOLUMES
CALCULATE CAPITAL AND O&M COSTS
SUM REGIONAL COSTS
SUM NATIONAL COSTS
LEGEND:
"IDEALIZED" FACILITIES
ACTUAL CASE STUDY SITES
INVENTORY OF POWER PLANTS > 25 MW

This procedure has been conducted for three scenarios under which coal combustion wastes might be disposed in order to "bracket" the probable cost impact of the proposed RCRA regulations on the coal-fired power generating industry. The scenarios were:

1. Representative Current Disposal Practices

   Wet disposal - fly ash and bottom ash are sluiced to a lagoon where the ash settles and eventually fills the disposal site, which is then closed, or the ash is later removed for ultimate disposal in a landfill.

   Dry disposal - Bottom ash is sluiced to a dewatering bin where the ash settles and "dry" ash is then disposed in a landfill.

Scrubber sludge was usually dewatered in a thickener and then, either separately or in combination with ash, might be trucked or pumped to a disposal site. Additional dewatering might also be accomplished by centrifugation or vacuum filtration. Stabilization agents might sometimes be used. Lagooning or landfilling was practiced depending upon the solids content of the disposed material.

Results from the state survey indicated the most common disposal method to be co-disposal of fly ash and bottom ash by wet sluicing to a lagoon. Effluent quality was controlled by National Pollutant Discharge Elimination System (NPDES) programs. Since most sites had not been inventoried, very little information concerning capacity of existing coal-fired utility waste disposal sites was available. Of peripheral interest was the fact that in most states sanitary landfill regulations were the governing factor for utility waste disposal.

Actual designs and current operating costs of the existing disposal methods for the powerplants in the case study assessment were not used due to wide variations in disposal practices and age of treatment and disposal facilities. In effect, the current practice disposal costs presented in the Interim Report are costs in 1979 dollars for building a disposal facility using pre-RCRA conventional technology.

Nonhazardous Disposal

2. Nonhazardous Disposal - as specified in RCRA Sections 1008 and 4004. The proposed guidelines for implementing RCRA Sections 1008 and 4004 formed the basis for determining costs for nonhazardous waste disposal. Criteria for site selection, design, leachate control, gas control, runoff control, operation and monitoring of disposal sites have been considered in development of preliminary designs and associated costs. The criteria were evaluated as recommended practices, but may be considered as requirements if states adopt the practices as rules. Nonhazardous waste disposal criteria (1008 and 4004) included siting facilities away from environmentally sensitive areas, protection

of surface water and groundwater through leachate and runoff control, and maintenance of daily operating conditions for a sanitary landfill.

Hazardous Disposal

3. Hazardous Disposal - as specified in RCRA Section 3004. The following partial list of criteria were among those considered in the development of cost impacts:

- Security
- Liner Systems
- Groundwater Monitoring
- Runoff/Leachate Collection/Treatment
- Buffer Zone
- Number of Waste Containment Cells
- Transportation/Transport Distance
- Diversion Structures
- Closure
- Post Closure Monitoring
- Liability Insurance
- Real Estate Cost
- Interest Rate
- Topography
- Sole Source Aquifer Zone
- Active Fault Zone
- Flood Plain
- Wetland
- Area of Sensitive Species
- Soils Data

A concept of "idealized designs" was developed, based on strict interpretation of the RCRA Proposed Rules and Guidelines for solid waste produced by conventional combustion. This was done to establish:

1. The lowest probable cost of compliance for a disposal site of a given size, and

2. The effects of economies of scale on waste disposal costs.

The idealized designs consisted of a variety of design assumptions which would correspond to a site which was virtually ideal for waste disposal. This included such factors as clay availability, climate, population density, etc. Four idealized designs were developed for each waste disposal scenario (identified above) based on waste disposal volumes of 800,000 to 60,000,000 cubic feet per year. (This reflects the range of coal-fired electric generating facilities in the nation.) The economies of scale were thus determined by varying only the disposal volume, while maintaining the same assumptions for all other variables. The cost data generated from the 16 idealized designs through regression analysis were then used to formulate mathematical models of capital and O&M costs as a function of waste disposal volume.

The cost equations or models that were developed to most closely correlate with the cost relationships (correlation coefficient of 0.98) were slope/intercept models in which the "idealized design" provides the scale factor or slope of a cost curve and the integration of real site data provides the intercepts. Capital cost in the equation is a log log ($y=ax^b$) function while the O&M cost is arthmetic ($y=ax+b$). As indicated earlier, this estimating procedure takes into account economies of scale.

One of the key elements of the cost estimating approach has been the selection of case study sites (24 sites total planned) that represent the range of factors which impact utility waste disposal costs (e.g., geographical locations, generating capacity, coal characteristics, effluent environmental control technology used, solid waste disposal technology, site characteristics, and potential siting restrictions). Six case study sites have been (18 more sites will be) visited and the following information collected:

1. Sources of Coal
    - Supplier
    - Quantity
    - District Number
    - Seam
    - General Characteristics
2. Boiler and Plant Operating Characteristics
    - Capacity
    - Age
    - Capacity Factors
    - Current Coal Burn
    - Projected 1985 Coal Burn
    - Emission Control Technology
3. Waste Disposal
    - Current Waste Production (Fly Ash, Bottom Ash, Scrubber Sludge, Water Treatment Sludge and Other Misc. Wastes)
    - Projected 1985 Waste Production
    - Current Waste Disposal Cost
    - Projected 1985 Disposal Cost
4. Schematic Flow Diagram of Facility
5. Site Characteristics
    - Topography
    - Soils, Geologic, and Climatic Data
    - Potential Disposal Site Locations and Configurations

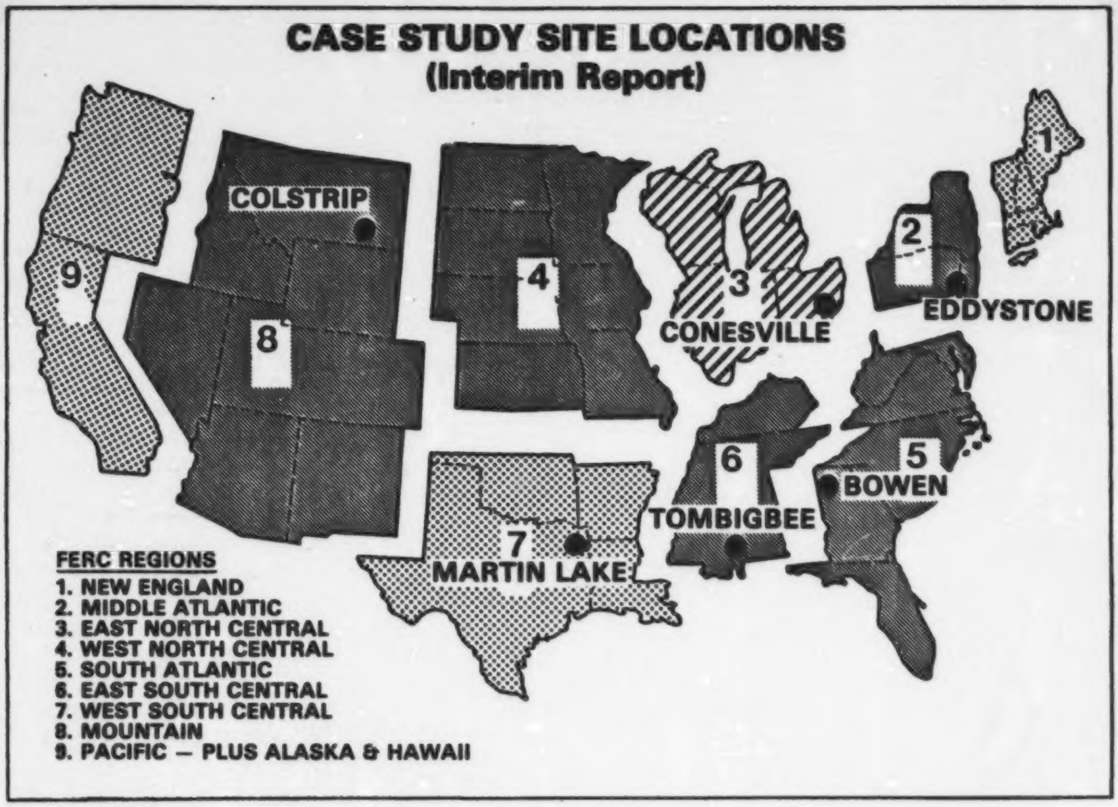
    Site specific cost estimates were made for transportation and treatment and disposal of coal combustion wastes generated.

Preliminary designs and cost estimates were prepared for each case study site based on the above data and the idealized or conceptual designs for each waste disposal scenario. The site-specific costs were then integrated into the equations based on the idealized designs to adjust the mathematical cost models to include various regional and site-specific factors.

To assess the potential range of costs that could be anticipated by construction of hazardous waste disposal facilities in environmentally sensitive areas (as defined by Section 3004 Propopsed Regulations), a representative site was selected near the Eddystone (Pennsylvania) plant. This site was located within a 500-year floodplain and a wetland area. Preliminary designs and cost estimates were prepared for this site based upon a range of possible construction methods that might be required.

Figure 5 shows the locations of the case study sites completed to date.

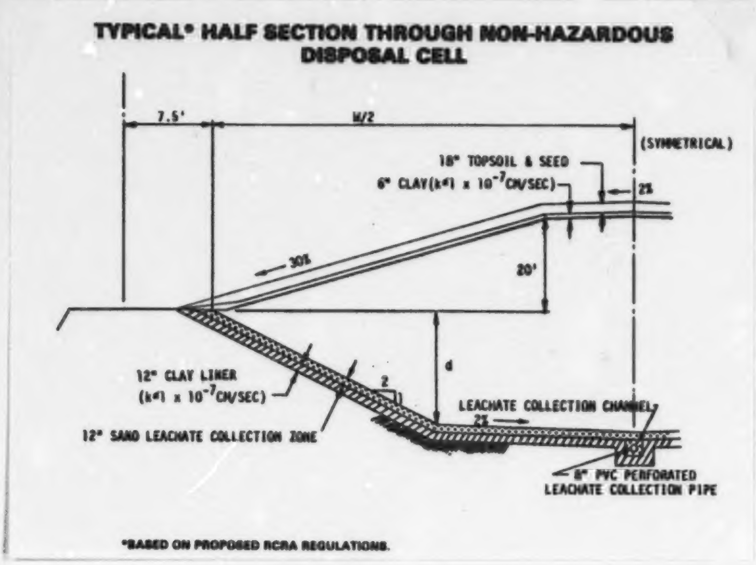
Figure 5


CASE STUDY SITE LOCATIONS
(Interim Report)
COLSTRIP
EDDYSTONE
CONESVILLE
BOWEN
TOMBIGBEE
MARTIN LAKE
1
2
3
4
5
6
7
8
9
FERC REGIONS
1. NEW ENGLAND
2. MIDDLE ATLANTIC
3. EAST NORTH CENTRAL
4. WEST NORTH CENTRAL
5. SOUTH ATLANTIC
6. EAST SOUTH CENTRAL
7. WEST SOUTH CENTRAL
8. MOUNTAIN
9. PACIFIC – PLUS ALASKA & HAWAII

Throughout the entire activity of developing site-specific preliminary engineering designs for disposal facilities, the primary objective in locating the disposal sites was to provide the most realistic cost-effective solid waste disposal system which would conform to the appropriate RCRA proposed Guideline. Sites were located to minimize haul distances, piping for leachate collection and treatment, and disposal cell construction costs.

Figures 6 and 7 below illustrate typical disposal cell design requirements to conform to RCRA Nonhazardous (Section 1008) and Hazardous (Section 3004) criteria, respectively. Each nonhazardous cell has a two-year capacity and active life while the hazardous cells have a one-year capacity and active life with a 6" daily cover.

Figure 6

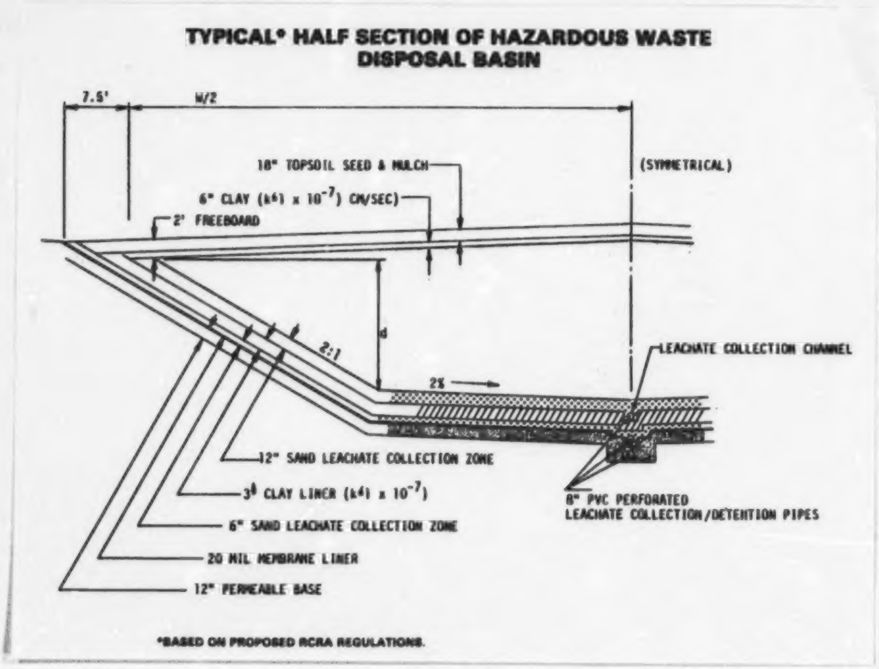


TYPICAL* HALF SECTION THROUGH NON-HAZARDOUS DISPOSAL CELL

7.5'
W/2
(SYMMETRICAL)
18" TOPSOIL & SEED
6" CLAY(k≤1 x 10^{-7}CM/SEC)
2%
30%
20'
d
12" CLAY LINER (k≤1 x 10^{-7}CM/SEC)
2
1
12" SAND LEACHATE COLLECTION ZONE
LEACHATE COLLECTION CHANNEL
2%
8" PVC PERFORATED LEACHATE COLLECTION PIPE

*BASED ON PROPOSED RCRA REGULATIONS.

Figure 7



TYPICAL* HALF SECTION OF HAZARDOUS WASTE DISPOSAL BASIN

7.5'
W/2
18" TOPSOIL SEED & MULCH
(SYMMETRICAL)
6" CLAY (k≤1 x 10^{-7}) CM/SEC)
2' FREEBOARD
d
2:1
2%
LEACHATE COLLECTION CHANNEL
12" SAND LEACHATE COLLECTION ZONE
3' CLAY LINER (k≤1 x 10^{-7})
6" SAND LEACHATE COLLECTION ZONE
20 MIL MEMBRANE LINER
12" PERMEABLE BASE
8" PVC PERFORATED LEACHATE COLLECTION/DETENTION PIPES

*BASED ON PROPOSED RCRA REGULATIONS.

Having developed the cost estimating models, the next step was to use these models to estimate regional and national cost impacts. This was done by recording the coal burned by each coal-fired power plant with greater than 25 MW capacity. An estimate of waste volume was then made based on the coal characteristics, the type of SO_2 scrubber employed (if any) and the method of disposing of the waste material. The waste volumes were entered into the cost models to obtain capital and O&M cost estimates for each of the approximately 400 plants in the U. S. with greater than 25 MW capacity. Regional costs were estimated by summing the disposal costs for all plants within a given Federal Energy Regulatory Commission (FERC) Region and making appropriate adjustments based on projected modifications to current practice. National costs were developed by summing the regional costs estimates. The results of these calculations are estimates of waste disposal costs for:

(1) current practice (both wet and dry disposal),

(2) disposal as a nonhazardous waste (RCRA Sections 1008 and 4004) and

(3) disposal as a hazardous waste (RCRA Section 3004).

Inherent in the above approach were the assumptions that (1) the disposal costs for each powerplant included in the case study assessment were representative of the costs for all plants in that region, and (2) weighted average disposal costs for the six plants were representative of the costs on a nationwide basis. As indicated earlier, extensive efforts were made to select case study sites that were representative of the generating plants in each Federal Energy Regulation Commission Region. It should be realized that this limited number of sites does not represent a statistically significant sampling of the total, in view of the number and magnitude of site-specific factors affecting costs. As a result, the cost estimates should be considered as preliminary, especially the regional cost estimates. However, they do have the advantage of being realistic estimates based on practical engineering designs. This limitation should be overcome in the remainder of the project by the increased number of site-specific estimates that will be used to adjust the cost curves developed from idealized designs.

The coal burn information used in the cost models was obtained primarily from the National Coal Association. Coal to be burned in 1985 was estimated by combining coal burn for existing facilities with the coal burn for the projected new facilities. The existing facilities were assumed to use 20% less coal in 1985 due to retirements and lower loads in the older units; the new facilities' coal requirements were calculated using the basic assumptions outlined above, except that 1977 state average BTU values were used.

The waste volume estimates used in the cost models were based on reported coal ash content values and calculations of projected scrubber sludge volumes. State-by-state coal ash contents were derived from Federal Power Commission report data representing average coal delivered in 1977. Using these ash content values, regional ash production rates were calculated for 1977.

Similarly, the ash content values were multiplied by the projected 1985 coal burn within each state and then combined to obtain the projected ash (fly ash, bottom ash, and boiler slag) for each region.

All scrubbers operating in 1977 were tabulated including capacity, percent sulfur, BTU/lb of coal, and removal efficiency. The dry weight of SO_2 sludge that would be produced assuming reliable scrubber operation (the 1977 "Sludge Potential") was then calculated on a plant-by-plant basis using the following formula:

$$SO_2 \text{ Sludge (tons} \times 10^3) = \frac{1.124 \times 10^5 \times \text{MW} \times \% \text{ S} \times \% \text{ Removal}}{\text{BTU/lb}}$$

This formula assumes continuous scrubber operation (except during boiler shutdown), "standard" load factors and heat rates. These figures were multiplied by 2 to adjust to 50% solids, and were then summed over the individual FERC Regions.

1985 projections for new facilities were obtained by applying the above formula to the coal to be burned in new facilities on a state-by-state basis. The 1977 state average values for percent sulfur and heat content of coal were used. A standard removal efficiency of 85% was assumed (lower percentages that may apply to low sulfur coal installations would have little effect on the quantities), and it was also assumed that 80% of all new facilities would be scrubbed. For example, the amount of sludge from new facilities in Pennsylvania was estimated as follows:

$$SO_2 \text{ Sludge} = \frac{1.124 \times 1450 \text{ MW} \times 2\% \text{ S} \times 85\% \text{ removal} \times 80\%}{11{,}897 \text{ BTU/lb}} = 196 \times 10^3 \text{ tons}$$

The 1985 state-by-state values were summed over each of the FERC Regions and added to the 1977 regional values. No reduction in 1977 sludge potential was assumed, since the scrubbers are on the newer units. While not specifically included for consideration in the original scope of work, the contractor, nevertheless, gave consideration to other related coal wastes (e.g., coal preparation wastes, boiler cleaning and blowdown wastes and coal pile drainage). Inter-relationships exist between coal preparation wastes and coal combustion residues, in that coal preparation techniques tend to concentrate the trace elements (as well as the sulfur and ash) in the coal wastes. Thus, the extent of coal preparation utilized will directly affect both the quantities and toxicity characteristics of the combustion residues. Conversely, the coal preparation wastes must be considered as a large volume with potentially significant environmental impact. Approximately three billion tons of coal preparation wastes have been accummulated in the United States, and 3,000 to 5,000 refuse dumps are estimated to exist. The current annual production of the 100 million tons of refuse is expected to reach 200 million tons within the next decade.

The Utility Solid Waste Activity Group (USWAG) has estimated the quantities of certain waste streams associated with typical coal-fired powerplants. Based on the USWAG estimates, and assuming a solids content of 25 percent, the generation rates for these wastes were estimated to be 121,000 to 1,750,000 tons in 1977, and are projected to increase to the range of 184,000 to 2,660,000 tons per year by 1985. These estimated waste tonnages suggest that the solid residues from treatment of the noncombustion related wastewater will contribute approximately 0.2 to 2.5% to the quantity of combustion related coal-fired utility wastes in 1985.

A typical disposal cost curve (Eddystone Station) illustrating capital costs for hazardous waste disposal is represented by Figure 8. Figure 9 illustrates a typical computation of regional disposal costs.

Figure 8

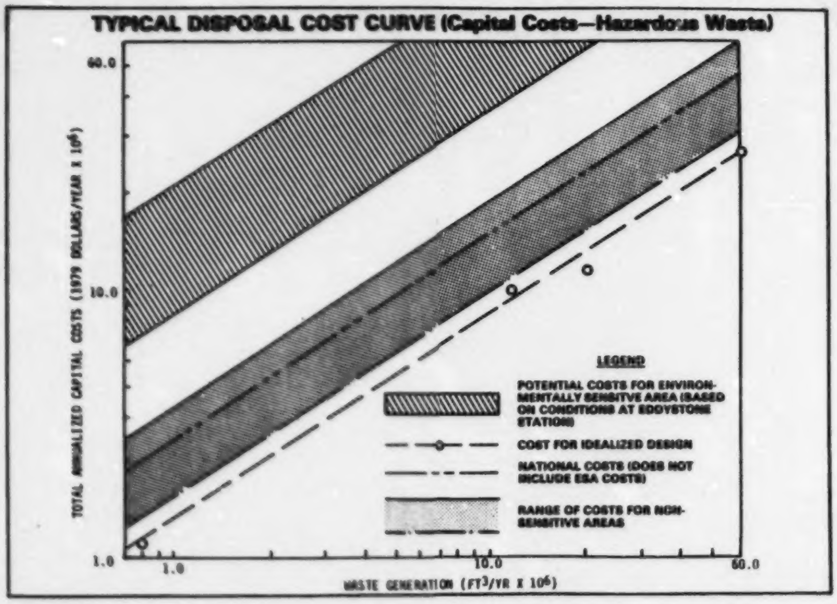
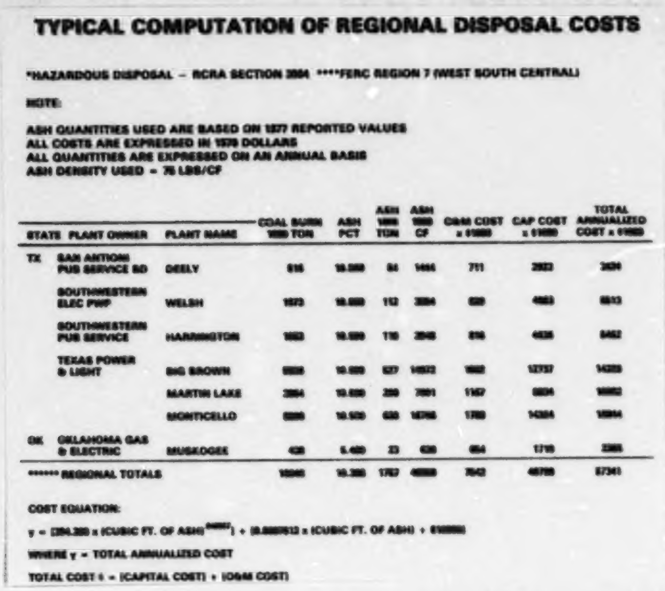


Figure 9



TYPICAL COMPUTATION OF REGIONAL DISPOSAL COSTS

*HAZARDOUS DISPOSAL – RCRA SECTION [illegible] ****FERC REGION 7 (WEST SOUTH CENTRAL)

NOTE:

ASH QUANTITIES USED ARE BASED ON 1977 REPORTED VALUES
ALL COSTS ARE EXPRESSED IN 1979 DOLLARS
ALL QUANTITIES ARE EXPRESSED ON AN ANNUAL BASIS
ASH DENSITY USED = 75 LBS/CF

STATE	PLANT OWNER	PLANT NAME	COAL BURN 1000 TON	ASH PCT	ASH 1000 TON	ASH 1000 CF	O&M COST x $1000	CAP COST x $1000	TOTAL ANNUALIZED COST x $1000
TX	SAN ANTONIO PUB SERVICE BD	DEELY	616	[illegible]	[illegible]	[illegible]	711	[illegible]	[illegible]
	SOUTHWESTERN ELEC PWP	WELSH	1573	[illegible]	112	[illegible]	[illegible]	[illegible]	[illegible]
	SOUTHWESTERN PUB SERVICE	HARRINGTON	[illegible]	[illegible]	110	[illegible]	[illegible]	[illegible]	[illegible]
	TEXAS POWER & LIGHT	BIG BROWN	[illegible]	[illegible]	627	[illegible]	[illegible]	12737	[illegible]
		MARTIN LAKE	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
		MONTICELLO	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
OK	OKLAHOMA GAS & ELECTRIC	MUSKOGEE	[illegible]	5.400	23	[illegible]	[illegible]	1719	[illegible]
****** REGIONAL TOTALS			[illegible]	[illegible]	1767	[illegible]	[illegible]	[illegible]	57341

COST EQUATION:

y = ([illegible] x (CUBIC FT. OF ASH)[illegible]) + ([illegible] x (CUBIC FT. OF ASH) + [illegible])

WHERE y = TOTAL ANNUALIZED COST

TOTAL COST $ = (CAPITAL COST) + (O&M COST)

A major factor affecting the total quantity of solid waste disposed was the amount of coal combustion by-products recovered for other uses. Likewise, modifications to certain operational aspects of the coal combustion power-generation-waste management cycle had the potential to reduce waste disposal costs.

According to the National Ash Association, approximately 17 milion tons of coal ash were utilized for a variety of industrial purposes in 1978, corresponding to over 24% of the 68 million tons produced.

One of the more significant potential areas currently in the development phase is metal recovery from coal ash. Results from a DOE-Ames study indicated that approximately 70% of the nation's need for aluminum could potentially be satisfied by chemical extraction from coal ash.

EPA representatives, during the comment period on the proposed Section 3001 classification system, indicated that most of the uses for coal ash would be considered "use constituting disposal." Consequently, ashes that were intended for these uses would have to be proven "nonhazardous" before their use would be allowed. It has been estimated that this interpretation of the RCRA regulations could effectively eliminate approximately 80% of coal ash utilization projected for 1985. Public concern over the use of materials similar to those classified as hazardous could reduce utilization even further. The net effort would include not only the loss of commercial value but induce a disposal cost as well. In 1985, the total cost in 1979 dollars, if a swing in value of $25 per ton were realized, would be approximately $720 million for that year.

While not examined in detail in the Interim Report, the continuing study will explore a number of processes which are being developed as a means of reducing waste disposal impacts. These include:

- combustion modification
- process modification
- waste treatment
- waste reuse
- resource recovery

These processes will be evaluated to assess their potential for ameliorating waste disposal costs by (1) reducing the volume of waste to be disposed and/or (2) altering the characteristics of the waste in order to reduce the unit cost of disposal.

Phase II

A portion of the Phase I - Interim Report - sought to address the probable impacts of the proposed RCRA Regulations on a number of emerging coal technologies. Generally, insufficient waste characterization data was available to determine how the wastes from most emerging technologies would be classified under RCRA. Therefore, preliminary waste generation rates were calculated and anticipated disposal costs for a hypothetical 1000 megawatt (MW) facility utilizing each emerging technology were developed in much the same manner employed for the case study sites.

The Phase II objective of the continuing study is to assess the implications of RCRA on an expanded number of emerging coal technologies involving advanced combustion, coal conversion and advanced environmental control processes.

Unlike the Phase I utility case studies, many of the emerging technology sites to be visited are pilot rather than full scale operations. This means that definitive, steady-state operating data will generally not be available because of the scale differences and the ongoing variability of operating conditions that are typical of pilot facilities. Therefore, the objective of the Phase II case study visits will be to obtain enough information to project the probable waste quantities and characteristics of similar full-scale units.

The number, size and location of commercial scale facilities utilizing the emerging technologies studied cannot be accurately determined at this time. Cost estimates made on a plant-by-plant basis (such as those in Phase I) are therefore not possible. Instead, scenarios will have to be developed for groups of hypothetical full-scale facilities based on process data obtained from the emerging technology case studies and disposal site data obtained from the utility case studies. Waste disposal costs will then be estimated using the cost models developed in Phase I.

Because the different emerging technologies affect waste generation in significantly different ways, some of these technologies are likely to be much more heavily impacted by RCRA than others. This project, therefore, will include a comparison of the relative waste disposal costs for various competing technologies in order to assess the probable impact of RCRA on their commercialization. Additionally, the projected commercialization rates of various emerging technologies will be factored into estimated costs of waste diposal for conventional (Phase I) powerplants.

A comparison of relative projected costs for disposal of wastes from emerging coal technology facilities is illustrated below under the discussion of the preliminary findings of the Interim Report.

Phase III

The Phase III activity, yet to be undertaken, is illustrated as the bottom line of Figure 1 above. The primary objective of Phase III will be to evaluate the results of Phases I and II on a nationwide basis through evaluation and assessment of financial, economic, geo-political and institutional impacts resulting from RCRA and other related Federal/environmental regulations so that the DOE can formulate and implement policies consistent with the NEP while recognizing legitimate environmental requirements.

Preliminary Findings and Conclusions

The data and numerical impacts which are summarized below come from the November 1979 Interim Report on the Utility Sector. While the numbers are of a preliminary nature at this time, it is, nevertheless, our belief that they may reflect a potential magnitude of RCRA impact which can be anticipated, if Sections of the RCRA Regulations applying to characterization and management of utility coal combustion derived wastes are promulgated 24 months from the present reporting and do not vary significantly from the existing proposed Sections 3001, 4004, 1008 and 3004.

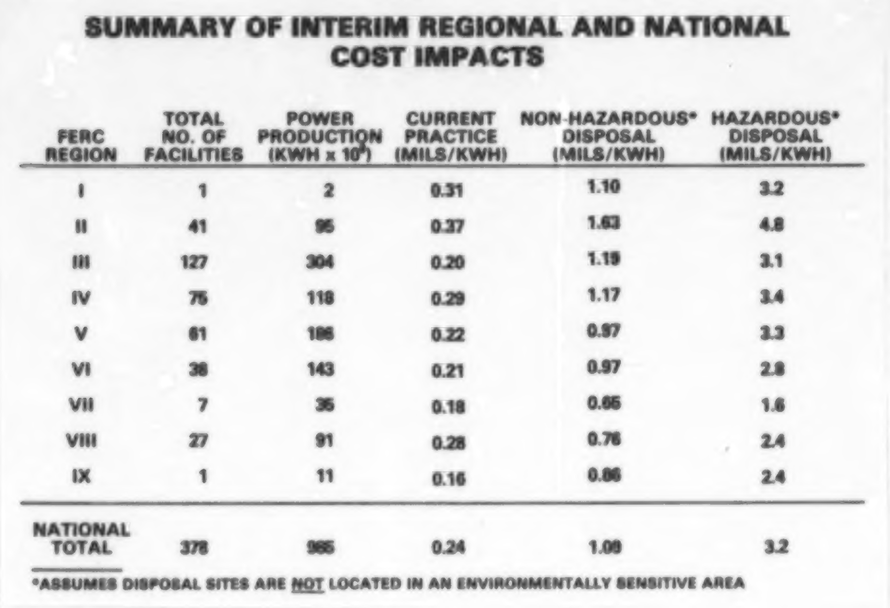
Relative to the six specific case study sites, analyses of the estimated total annual disposal costs indicated the following annual incremental cost increases for nonhazardous and hazardous waste disposal as compared to estimated current disposal costs:

Plant	Waste Generation 1985 Tons/Year	Nonhzardous Increment	Hazardous Increment
Bowen	1,069,000	777%	2682%
Conesville	802,000	464%	1334%
Tombigbee	160,000	438%	1520%
Eddystone	116,000	242%	913%
Martin Lake	1,557,000	225%	713%
Colstrip	1,148,000	334%	1147%

In the above tabulation, total annual expenditure would be largely a function of the amount of waste generated.

As explained earlier, regional and national costs were computed based on curve equations for capital and O&M costs for each case study site in a region (extrapolated later to all facilities greater the 25MW capacity within a given region), and for each waste disposal scenario considered. Summation of regional impacts provided an estimate of national cost impacts. Figure 10 below summarizes these results.

Figure 10



SUMMARY OF INTERIM REGIONAL AND NATIONAL COST IMPACTS

FERC REGION	TOTAL NO. OF FACILITIES	POWER PRODUCTION (KWH x 10^9)	CURRENT PRACTICE (MILS/KWH)	NON-HAZARDOUS* DISPOSAL (MILS/KWH)	HAZARDOUS* DISPOSAL (MILS/KWH)
I	1	2	0.31	1.10	3.2
II	41	95	0.37	1.63	4.8
III	127	304	0.20	1.19	3.1
IV	75	118	0.29	1.17	3.4
V	61	186	0.22	0.97	3.3
VI	38	143	0.21	0.97	2.8
VII	7	35	0.18	0.65	1.6
VIII	27	91	0.28	0.76	2.4
IX	1	11	0.16	0.86	2.4
NATIONAL TOTAL	378	985	0.24	1.09	3.2

*ASSUMES DISPOSAL SITES ARE NOT LOCATED IN AN ENVIRONMENTALLY SENSITIVE AREA

In terms of the incremental impacts, the additional national average cost per dry ton of ash disposed was estimated to be $13.81 (assuming that RCRA 4004 type facilities were required instead of current practice), and $47.54 (assuming that RCRA 3004 type facilities were required instead of current practice).

On the average, preliminary national costs for waste disposal, based upon direct costs alone under RCRA, potentially reflected increased costs from 4 to 13 times greater than for the costs for conventional disposal practices, depending upon whether the wastes were classified as nonhazardous. Further, the total cost for hazardous waste disposal facilities located in environmentally sensitive areas (such as wetlands and floodplains) were expected to be from two to five times greater than for similar facilities located in nonsensitive areas based on specific design considerations for one particular site. (This will be further investigated in the continuing work.)

On an annual basis, capital costs comprised 80 to 88% of the total disposal costs for implementation of RCRA Section 4004, nonhazardous wastes. Associated O&M accounted for the balance (12 to 20%).

Capital costs comprised 88 to 99% of the total disposal costs for RCRA Section 3004, Hazardous Wastes. The liner system and earthwork represented the major capital cost components. Waste transport was the major O&M cost.

In most cases, the contractor found that current practices employed for disposal of coal combustion wastes would not conform to RCRA, irrespective of whether the wastes could be declared to be hazardous.

A potentially ameliorating influence on the cost of coal combustion waste disposal could be the cost offsets attributed to waste utilization techniques. The EPA proposed definition of "other discarded material" to include reused material "if such use constitutes disposal" could include traditional uses of coal-combustion by-products such as: filler or aggregate in bituminous concrete, stabilized roadbase compositions, blast grit, roofing granules, highway ice control, structural embankments, and fill for land improvement. Contrary to the original intent of the Act, the proposed RCRA regulations could severely limit existing methods of coal combustion by-product utilization.

Early indications regarding the RCRA potential for impact of the emerging coal technologies under development or supported by DOE/OCU showed that relative costs for disposal of residues from various emerging coal technologies could vary considerably from one technology to another. These cost differences appeared to be of sufficient magnitude to affect the rates of commercialization of various competing technologies.

A comparative summary of waste disposal costs of several of these emerging technologies with conventional coal combustion waste costs is presented in Figure 11 below.

Figure 11

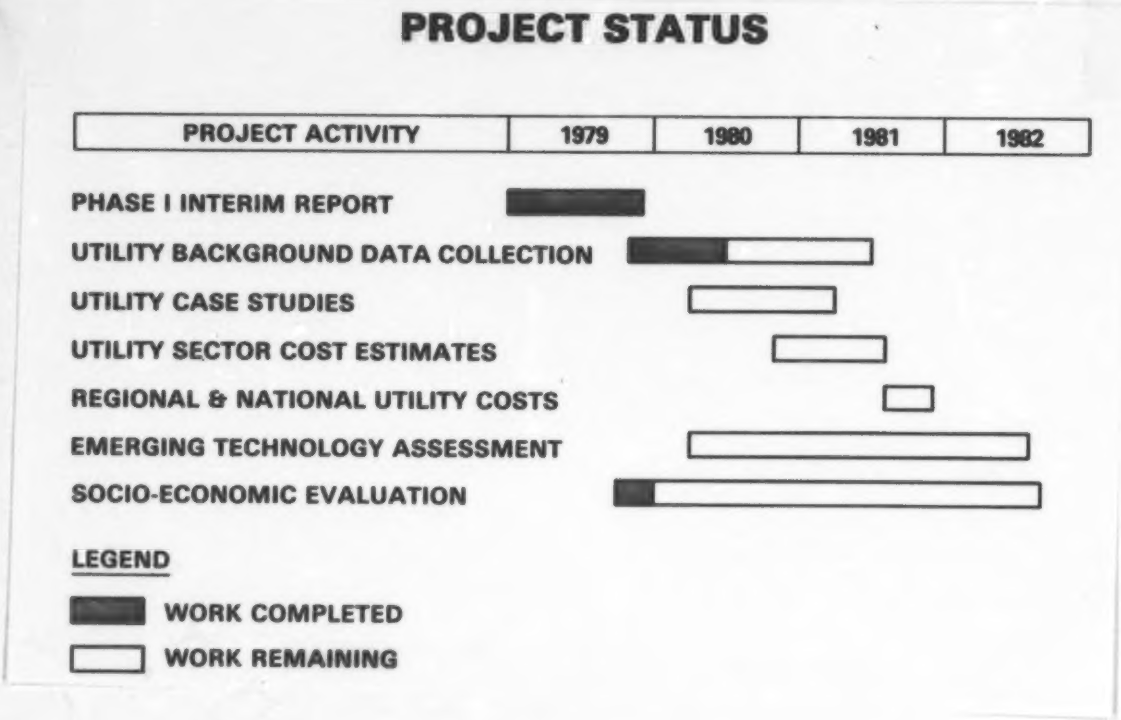
PROJECTED WASTE DISPOSAL COSTS FOR EMERGING COAL TECHNOLOGIES*

	WASTE DISPOSAL COST (1979 mils/kwh)		
POWER GENERATION TECHNOLOGY	EXISTING PRACTICE	NON-HAZARDOUS DISPOSAL	HAZARDOUS DISPOSAL
CONVENTIONAL W/FGD			
250 MW	0.57	2.30	6.83
500 MW	0.36	1.80	5.25
1000 MW	0.27	1.43	4.07
CONVENTIONAL W/O FGD			
250 MW	0.35	1.46	4.37
500 MW	0.25	1.12	3.33
1000 MW	0.18	0.87	2.13
SRC II LIQUEFACTION			
250 MW	0.14	0.77	2.12
500 MW	0.13	0.76	2.06
1000 MW	0.13	0.74	2.03
LOW-BTU GASIFICATION			
250 MW	—	—	—
500 MW	—	—	—
1000 MW	0.16	0.79	2.29
ATMOSPHERIC FLUIDIZED-BED			
250 MW	0.58	2.77	8.19
500 MW	0.42	2.18	6.31
1000 MW	0.32	1.74	4.90
PRESSURIZED FLUIDIZED-BED			
250 MW	0.40	2.24	6.68
500 MW	0.35	1.75	5.13
1000 MW	0.26	1.39	3.97

*INTERIM REPORT, PHASE I - UTILITY SECTOR (NOVEMBER, 1979)

Figure 12 reflects the current project status.

Figure 12



The author wishes to acknowledge the outstanding performance of his contractor, Engineering-Science, in the development of the preliminary data for the Interim Report on the Utility Sector and the results which are reported in this paper.

EPRI FGD SLUDGE DISPOSAL DEMONSTRATION AND SITE MONITORING PROJECTS

by

Dean M. Golden, P.E.*

ABSTRACT

The increasingly stringent air quality requirements on power plants necessitated the clean-up of flue gas stack emissions, resulting in ever-increasing quantities of solid wastes from the combustion of coal. Presently, the most used type of flue gas desulfurization (FGD) scrubber is the wet scrubber. This kind of scrubber is expected to dominate the industry into the 1990's. The waste produced by wet scrubbing is a sludge composed of the scrubbing liquor, calcium sulfite/ sulfate solids and varying quantities of flyash. EPRI has a number of research projects recently completed or underway exploring new options for disposal of the FGD sludges. This paper summarizes the results of two projects, one evaluating the stacking method for gypsum disposal, and the other a site monitoring investigation to assess the acceptability of sludge/ash fixation.

The results of the FGD gypsum stacking demonstration indicate that this oxidized FGD sludge has settling, dewatering, and structured characteristics similar to, and in some cases, superior to phosphate gypsum, making this a viable option for disposal.

The results of the site monitoring of the first full-scale fixed sludge disposal system are still inconclusive. Well monitoring indicates some impact on groundwater quality in the vicinity of the disposal facilities. Field permeabilities of the fixed sludge samples were found to be higher than laboratory test results.

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EPRI FGD Sludge Disposal Demonstration and Site Monitoring Projects

by

Dean M. Golden, P.E.

INTRODUCTION

The disposal of the millions of tons of coal ash and flue gas desulfurization (FGD) sludges produced each year in the United States by coal power plants is becoming a bigger problem each year. Nearly 70 million tons of coal ash were produced last year. The number of power plants with scrubbers is increasing rapidly. A total of 33 utilities are currently using scrubbers on 62 units, equivalent to 20,630 MW of installed capacity as of November 1, 1979. A total of 51 additional units are under construction, contract or letter of intent as of the same date, totalling an additional 24,385 MW of "scrubbed" capacity.[1] The waste produced by the wet scrubbers, which currently dominate the industry, is a sludge composed of scrubbing liquor, calcium sulfite/sulfate solids and varying quantities of fly ash.

At the present time 51 percent of ash disposal is done by sluicing in slurry form to disposal ponds and the remainder is handled dry and placed in landfills.[2] Currently 60 percent of the FGD sludges are ponded and the remainder disposed of dry in landfills.

The Electric Power Research Institute (EPRI) has a number of research projects recently completed or underway evaluating new disposal options for the coal combustion by-products. In order to keep this paper to reasonable length, only two projects will be reviewed, one evaluating the stacking method for gypsum disposal and the other a site monitoring investigation to assess the acceptability of sludge fixation using fly ash.

SLUDGE FIXATION PROCESS EVALUATION

Stabilization or fixation processes based on pozzolanic reactions (also called lime-based reactions) are being utilized by an increasing number of utilities (five as of November 1, 1979[1]) to reduce the environmental impact of these materials by altering their physical and chemical properties. Stabilized FGD sludge has been landfilled at the Conesville Generating Station of Columbus and Southern Ohio Electric Company (CSOE) since January 1977. This facility was the first full-scale application of the commercial sludge fixation process marketed by IU Conversion Systems, Inc. (IUCS). Since this appeared to be a promising alternative to disposal by ponding EPRI initiated a project (RP1406) to evaluate the facility over a three year period to determine the success of the fixation process. The principal objectives of the EPRI research project which evolved are:

- to determine if the actual stabilized sludge produced under full-scale conditions compare satisfactorily with laboratory and pilot scale observations
- to determine if the method of disposal as practiced at Conesville is environmentally acceptable (that is to determine if there are detrimental leachate, runoff, or future land use problems associated with the disposal option)
- to determine what operating problems, if any, the sludge disposal method causes for the utility
- to determine if the method of disposal will satisfy current and projected regulatory requirements
- to develop and verify a predictive groundwater flow model with generic applicability to similar sites

This research project involving Michael Baker, Jr. (Baker) and Battelle, Pacific Northwest Laboratories (Battelle) as the EPRI contractors, is being conducted in two phases. Phase I work conducted by Baker was completed in September 1979 and consisted primarily of the data gathering activities necessary for an understanding of the site specific disposal operations and background hydrogeologic conditions, the initiation of groundwater monitoring, and planning for the more extensive Phase II monitoring. The Phase I preliminary findings are detailed in EPRI Report FP-1172 entitled Monitoring the Fixed Sludge Landfill, Conesville, Ohio - Phase I. A companion Phase I study was completed in March 1980 by Battelle in which the background geohydrologic and site data collected largely by Baker was used to develop and preliminarily verify a predictive model for the groundwater flow system. The Phase I work by Battelle is detailed in EPRI Report CS-1355 entitled Modeling the Fixed FGD Sludge Landfill-Conesville, Ohio (Phase I).

Baker's Phase II activities involve three years of groundwater and runoff monitoring around the fixed sludge disposal site, annual physical testing and leachate analysis on samples of the fixed sludge, general observations of the disposal system operations, and final evaluations of the overall process. Battelle Phase II activities involve additional modeling work to develop a model users guide for generic application. This paper summarizes the Phase II work of Baker during the past year.

FIELD INVESTIGATIONS

Monitoring Wells

A monitoring network of 30 wells has been installed and sampled regularly for this project. Thirteen of the Phase I monitoring wells have been sampled since the beginning of the project. The 17 new monitoring wells are arranged in five clusters of three wells each and have been sampled during the Phase II monitoring period. Figure 1 illustrates the locations of all wells installed to date. Table 1 summarizes the basic well data including total depth, screened interval, and well nomenclature.

Geophysical Well Logging

After installing the 17 new monitoring wells in the Phase II work, geophysical logging was conducted in most of the wells through the polyvinyl chloride well casing. It was not possible to do the logging in uncased holes as is the usual practice, since the wells go through thick deposits of unconsolidated glacial outwash material so uncased wells would not remain open. The logging provided density, temperature, and fluid resistivity data to supplement the other known subsurface information. This would also provide another means of detecting groundwater contamination separate from the ongoing well water sampling program. Gamma logs were also made to identify clay deposits.

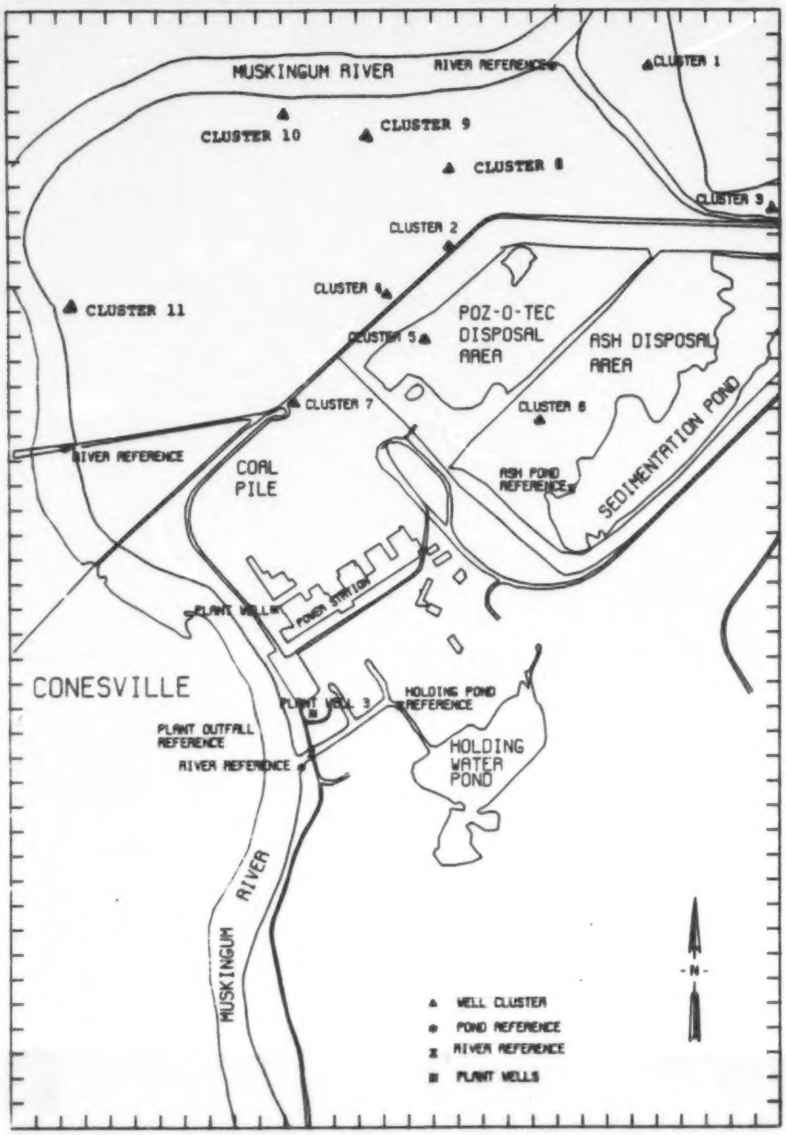
MUSKINGUM RIVER
RIVER REFERENCE
CLUSTER 1
CLUSTER 10
CLUSTER 9
CLUSTER 8
CLUSTER 3
CLUSTER 2
CLUSTER 4
CLUSTER 11
CLUSTER 5
POZ-O-TEC DISPOSAL AREA
ASH DISPOSAL AREA
CLUSTER 7
CLUSTER 6
SEDIMENTATION POND
RIVER REFERENCE
COAL PILE
ASH POND REFERENCE
POWER STATION
PLANT WELLS
CONESVILLE
PLANT WELL 3
HOLDING POND REFERENCE
PLANT OUTFALL REFERENCE
RIVER REFERENCE
HOLDING WATER POND
MUSKINGUM RIVER
N
▲ WELL CLUSTER
● POND REFERENCE
x RIVER REFERENCE
■ PLANT WELLS

Figure 1. Locations of Groundwater Monitoring Wells at the Conesville Station

Note cluster designation and well numbers below:

Cluster 1 - MB-1,2,3
Cluster 2 - MB-4,5
Cluster 3 - MB-6
Cluster 4 - MB-7,8,9
Cluster 5 - MB-10,11,30
Cluster 6 - MB-12,13,31
Cluster 7 - MB-14,15,16,17
Cluster 8 - MB-18,19,20
Cluster 9 - MB-21,22,23
Cluster 10 - MB-24,25,26
Cluster 11 - MB-27,28,29

TABLE 1

SUMMARY OF MONITORING WELL DETAILS

Well Numbers and Cluster Designation	Total Well Depth Below Ground	Interval of Slotted Well Casing
MB-1	49 ft (14.9 m)	29-49 ft (8.8-14.9 m)
MB-4	25 ft (7.6 m)	15-25 ft (4.6-7.6 m)
MB-5	49.5 ft (15.1 m)	34.5-49.5 ft (10.5-15.1 m)
MB-6	50 ft (15.3 m)	40-50 ft (12.2-15.3 m)
MB-7	25 ft (7.6 m)	15-25 ft (4.6-7.6 m)
MB-8	49.5 ft (15.1 m)	34.5-49.5 ft (10.5-15.1 m)
MB-9	90 ft (27.5 m)	75-90 ft (22.9-27.5 m)
MB-10	45.5 ft (13.9 m)[a]	35.5-45.5 ft (10.8-13.9 m)
MB-11	66.5 ft (20.3 m)[a]	56.5-66.5 ft (17.2-20.3 m)
MB-30	103 ft (31.4 m)	88-103 ft (26.8-31.4 m)
MB-12	38.5 ft (12 m)[a]	33.5-38.5 ft (10.2-11.7 m)
MB-13	58.5 ft (17.8 m)[a]	48.5-58.5 ft (15-17.8 m)
MB-31	107.5 ft (32.8 m)	92.5-107.5 ft (28.2-32.8 m)
MB-14	25 ft (7.6 m)	15-25 ft (4.6-7.6 m)
MB-15	50 ft (15.3 m)	35-50 ft (10.7-15.3 m)
MB-18	25.5 ft (7.8 m)	15.5-25.5 ft (4.7-7.8 m)
MB-19	45 ft (13.7 m)	35-45 ft (10.7-13.7 m)
MB-20	70 ft (21.4 m)	60-70 ft (18.3-21.4 m)
MB-21	25 ft (7.6 m)	15-25 ft (4.6-7.6 m)
MB-22	45 ft (13.7 m)	35-45 ft (10.7-13.7 m)
MB-23	94 ft (28.7 m)	79-94 ft (24.1-28.7 m)
MB-24	18.5 ft (5.6 m)	8.5-18.5 ft (2.6-5.6 m)
MB-25	39.5 ft (12.0 m)	29.5-39.5 ft (9.0-12.0 m)
MB-26	85 ft (25.9 m)	70-85 ft (21.4-25.9 m)
MB-27	30 ft (9.2 m)	20-30 ft (6.1-9.2 m)
MB-28	49 ft (14.9 m)	34-49 ft (10.4-14.9 m)
MB-29	70 ft (21.4 m)	55-70 ft (16.8-21.4 m)
MB-32	29 ft (8.8 m)	19-29 ft (5.8-8.8 m)
MB-33	49 ft (14.9 m)	39-49 ft (11.9-14.9 m)
MB-34	74.5 ft (22.7 m)	64.5-74.5 ft (19.7-22.7 m)

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[a] Casing added to well since Phase I due to continued filling activities.

Water Sampling and Field Measurements

Water samples were collected and tested for each monitoring well four times during the Phase II study, in December 1979, and March, May and July 1980. These results were compared to those obtained during Phase I.

Runoff samples from the fixed sludge disposal area were collected from two locations at the time of or shortly after storm events that occurred during the last three sampling periods. These runoff samples were collected for the purpose of evaluating the impact of leachate produced as a result of direct precipitation to determine if special handling other than temporary retention for suspended solids removal might be necessary.

The filtrate resulting from the vacuum filters used on the thickened FGD sludge has been continually discharged to the ash pond rather than being recycled to the scrubber system as originally planned by CSOE. The reported reasons are the unexpected high suspended solids levels remaining after filtration which caused premature scouring of the return piping. Samples of this filtrate were collected for analysis.

FIXED SLUDGE SAMPLING AND TESTING

In November 1979 four different areas within the fixed sludge disposal area were sampled using standard drilling techniques to primarily obtain core samples of the stabilized material known as "Poz-O-Tec®," that were one year old or more. In addition, thin wall samples were collected using Shelby tubes of two week old material for physical testing. Physical testing was done to compare the results with laboratory and pilot scale studies. The tests conducted the stabilized sludge and the methods used (shown in parentheses) are as follows:

- Moisture Content (ASTM Standard Method D-2216 for soils)
- Unconfined Compressive Strength (ASTM D-2166)
- Permeability (As recommended in U.S. Army Corps of Engineers Testing Manual EM 110-6-1906)
- Wet Density

- Dry Density
- Triaxial Strength (ASTM D-2850)
- Leachability (Same procedure as permeability determination above. A head of 25 feet (7.6 m) used to permit adequate contact time of leaching medium with sample. Distilled water used as leaching medium. Standard of 2100 ml of leachate generated.)

In July 1980 a few additional samples were obtained during a sampling program conducted by IU Conversion Systems for comparison with results with replicate samples tested by IUCS.

OBSERVATION OF DISPOSAL OPERATIONS

As stated earlier, in the objective of this project, the disposal operations were to be observed to see if any operational difficulties arise from application of the IUCS sludge fixation process. This was done in conjunction with quarterly plant visits for monitoring and sample collection.

WATER SAMPLING AND GEOPHYSICAL LOGGING RESULTS

The water quality in the monitoring wells upstream of the disposal area, believed to be indicative of background conditions in the vicinity remained very consistent as compared to the Phase I observations.

Monitoring well data obtained from wells near the ash pond which is adjacent to the fixed sludge landfill show the greatest effect on the groundwater quality. Levels in excess of recommended EPA Secondary Drinking Water Standards for pH, total dissolved solids, sulfate, and total iron were noted in two wells located directly in the ash disposal area. Four wells down-gradient from the ash pond showed elevated levels for the same parameters, although somewhat less, which may be indicative of soil attenuation.

The most significant change in quality found by comparing Phase I with Phase II data were the levels of calcium, total dissolved solids, sulfite and magnesium observed in those wells in the ash pond. The changes in levels of these parameters is believed to be a result of the practice at Conesville of discharging filtrate from the vacuum filtration of the thickened FGD sludge, and/or the raw sludge discharged to the ash pond under plant upset conditions.

The Phase II sampling in two wells located near the emergency sludge pond showed improvement over the Phase I results. The higher Phase I levels is attributable to the initial unlined condition of the emergency sludge pond.

As in Phase I, the groundwater quality observed in the two wells situated within the fixed sludge landfill did not show leachate permeating through the fixed sludge material. On the other hand the two wells located to monitor the effect of surface runoff from the fixed sludge area, did show significant increases in calcium, conductivity, total dissolved solids, sulfate, and magnesium. Until July 1980, the runoff was collected in a small pond on the combined northwestern corner of the sludge disposal area. The underlying sand and gravel glacial outwash were easily infiltrated with the runoff. Water quality analyses on the runoff itself confirmed the water quality relationship with the well samples. The runoff collection pond was lined in July 1980 with fixed sludge material.

During Phase II, average fluid resistivities for most of the monitoring wells were determined from geophysical well logs. The resistivities measured were highest (indicating least ion-enriched water) north of the fixed sludge and ash pond areas and at the background monitoring wells. The lowest resistivities (indicating the most ion-enriched water) were found at the wells immediately down-gradient of the ash pond.

FIXED SLUDGE TEST RESULTS

As described earlier in this paper, an extensive laboratory testing program on samples of the fixed sludge was acquired at the Conesville site during the Phase II activity. The Conesville landfill has been sampled on other occasions by or for the process vendor, IU Conversion Systems. There are differences in the sampling procedures used by Baker compared to IUCS. As reported earlier Baker used standard geotechnical test procedures commonly used. Rather than using standardized core drilling and split spoon sampling, IUCS uses a more elaborate procedure of digging a test pit to excavate test samples. The test samples were later drilled from the excavated block samples under laboratory controlled conditions.

Physical Testing

Moisture content determinations were made on 120 samples. The majority of the moisture contents were found to lie within the range of 45 to 65 percent (by dry weight). No significant differences in moisture content were found between samples of different age. The optimum moisture content for compaction of the Poz-O-Tec® fixed sludge material has been reported to be in the range of 35 to 55 percent[1].

Bulk and dry density measurements were taken on 24 fixed sludge samples acquired by Baker during the Phase II study. A majority of the bulk densities measured were found to be between 1375 to 1450 kg/m^3 (85 to 90 pounds per cubic feet), while the dry densities ranged between 809 to 955 kg/m^3 (50-59 pcf). Both the bulk and dry densities determined by Baker are below the reported densities of the Poz-O-Tec® fixed sludge material. Reported bulk densities ranged from 1537 to 1618 kg/m^3 (95-100 pcf) while dry densities ranged from 1052-1375 kg/mn^3 (65-85 pcf).

Permeability

As was noted in the Phase I report[3], freshly placed Poz-O-Tec® fixed sludge materials are claimed to have permeabilities in the range of 10^{-6} cm/sec while material that has been cured for 14 to 28 days can develop permeabilities reaching 10^{-7} cm/sec. During Phase II, year 1, sixteen permeability determinations were attempted and three samples fell apart during trimming. Table 2 shows the permeabilities and the estimated age of the sample at the time of collection.

A review of the permeability data shows that 62% of the fixed sludge samples had permeabilities in the range of 10^{-4} to 10^{-5} cm/sec, which is the range associated with untreated FGD sludges and fly ash. Obviously, these permeability levels would jeopardize the usefulness of this Poz-O-Tec® fixed sludge material as a liner in disposal sites given that the recommended EPA criteria for non-hazardous wastes is 10^{-7} cm/sec.

TABLE 2

POZ-O-TEC® SAMPLE PERMEABILITIES[4]

Permeability Test Number	Sample Type	Poz-O-Tec Age At Time of Test	Boring Number	Sample Depth	Measured Permeability (cm/sec)	Remarks
P-3	Shelby	4.5 months	MBp-2	2.4 ft.	1.7×10^{-4}	Sample crumbly (poor cohesion)
P-5	Shelby	5 months	MBp-3	0.3 ft.	2.2×10^{-6}	
P-7	Core	+1.4 years	MBp-4	1 ft.	2.1×10^{-5}	
P-8	Core	+1.4 years	MBp-4	5.7 ft.	5.2×10^{-5}	Sample is relatively older than P-7
P-9	Core	+1.4 years	MBp-4	9.3 ft.	9.4×10^{-5}	Sample is relatively older than P-8
P-12	Core	+1.4 years	MBp-6	6.6 ft.	2.3×10^{-5}	Sample is relatively older than PP8 but younger than P-9
P-1	Core	+1.5 years	MBp-1	7.5 ft.	1.57×10^{-4}	Sample vuggy
P-15	Core	+1.5 years	MBp-10	1.4 ft	4.56×10^{-6}	
P-10	Core	+1.5 years	MBp-5	4.6 ft.	1.3×10^{-5}	Sample porosity 64%
P-4	Core	+1.5 years	MBp-2	5.7 ft.	8.3×10^{-6}	Sample porosity 67%
P-11	Core	+1.5 years	MBp-5	8.5 ft.	2.6×10^{-4}	Sample porosity 58%, sample is relatively older than P-10.
P-14	Core	+1.6 years	MBp-7	10.5 ft.	4.61×10^{-6}	
P-13	Core	+1.6 years	MBp.7	1.7 ft.	8.68×10^{-6}	

At the time of writing this paper, the results of the July 1980 permeability tests on the samples excavated by IUCS have not been completed. Tentative results indicate that the permeabilities are lower than those using the standard coring methods of sampling, generally in the range of 10^{-6} cm/sec. The wide disparity in the results of the two tests indicates the effect of sample collection methodology on permeability data.

Strength Tests

Compressive strength determinations on the undisturbed Poz-O-Tec® samples showed levels high enough to support normal foundation loads. A majority of the samples had strengths of about 7 kg/cm^2 (100 psi).

Leachate Generation

Although, a total of eight leachate generation tests were planned on undisturbed samples, only five tests could be completed since three samples fell apart during trimming. A triaxial compression chamber was used as a permeameter, with distilled water as the leaching medium. In order to facilitate a complete water quality analysis, 2100 m of leachate was produced from each sample, representing 17 to 42 pore volume replacements. Table 3 shows the range of results from the five samples and compares them to the Environmental Protection Agency drinking water standards.

DISPOSAL OPERATIONS

The Phase I monitoring report[3] discussed the problem in handling the overly wet fixed sludge at Conesville. This condition occurs because of the presence of magnesium in the lime that is used in the scrubbing process which increases the SO_2 removal efficiency. The magnesium content decreases the efficiency of sludge thickening and vacuum filtration. Although the high magnesium lime was originally specified in the design of the IUCS disposal system, its effect on the dewatering characteristics was not anticipated. The resulting wet filter cake can be further "dried" by the addition of more fly ash, however, only fly ash from Units 5 and 6 is available for sludge fixation. Fly ash from the other units is sluiced to the ash pond and is unavailable without major system changes. A just released EPRI report[5] found that fly ash contents of 75 percent or greater are required for good long term strength behavior,

TABLE 3

FIXED FGD SLUDGE LEACHATE CHARACTERISTICS
CONESVILLE POWER STATION

Constituent	Leachate Range of Results[a]	EPA[b] Standard for Drinking Water
pH (standard units)	6.4 to 8.5	6.5 to 8.5 (S)
Net Alkalinity	1 to 33	--
Sulfate	85 to 1350	250(S)
TDS	177 to 2184	500(S)
COD	<5 to 8	--
Total Iron	.02 to .06	0.3(S)
Calcium	67 to 604	--
Magnesium	22.3 to 40.6	--
Barium	<1 to .1	1(P)
Cadmium	<.01 to .01	.01(P)
Chromium	<.01 to .02	.05(P)
Lead	<.03 to .065	.05(P)
Silver	<.01 to .02	.05(P)
Arsenic	.005 to .016	.05(P)
Selenium	<.005 to .011	.01(P)
Mercury (μg/l)	<.5 to 2.0	2(P)
Conductance (μmhos @ 25°C)	272 to 2439	--
Sulfite	<1 to 8	--
Boron	<.5 to 1.6	--

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[a] Results in mg/l unless otherwise specified.

[b] (P) Mandatory Interim Primary Drinking Water Std; (S) Recommended Secondary Drinking Water Std.

therefore the addition of fly ash would be helpful both from a strength and a drying standpoint.

Because the Poz-O-Tec® fixed sludge is wetter than desirable, and no more ash available, IU Conversion Systems has recommended that the fresh material be allowed to cure in surge piles at the base of the radial-arm conveyor for approximately 3 to 6 days before handling. After this period of time IUCS believes that the material can be handled easily and placed in two-foot lifts. Theory often does not work out in practice. CSOE operators have found that the partially cured materials will not support their equipment during excavation from the piles and spreading. To reduce pumping beneath the CSOE dozer, the fixed sludge is being spread in thicker lifts. CSOE relies on the traffic of the dozer and the two 50-ton trucks used for disposal activities for compaction. The results of the Phase II testing show that there is insufficient compaction. Another contributing factor the disappointing permeability results is the over curing of the Poz-O-Tec® material in the surge piles from the radial-arm conveyors. This over curing creates hard "boulders" of fixed sludge material which result in bridging and voids when they are spread in the lifts. Therefore the higher than expected permeabilities of the Poz-O-Tec® material is probably a result of a combination of the thicker lifts, the large boulders in the fill, and insufficient compaction of the fixed sludge. Figures 2 and 3 show the boulders in the surge pile and the lifts.

COMPLIANCE WITH APPLICABLE REGULATIONS

Although the Phase II investigation shows that the local groundwater quality has been degraded to some degree, by the surface runoff discharges associated with the IUCS sludge fixation process, it is believed that applicable Federal waste disposal regulations are presently being met. The EPA Secondary Drinking Water Standards which have been exceeded are currently only proposed standards. The U.S. Interim Primary Drinking Water Standards which are now mandatory have not been violated. The runoff from the fixed sludge disposal area will meet the current NPDES requirements provided sedimentation ponds are provided to control suspended solids.

Under the 1980 Resource Conservation and Recovery Act Amendments (RCRA) now awaiting signature by President Carter, the fixed sludge material is exempted

Figure 2. "Boulders" in newly excavated Poz-O-Tec from the surge pile.

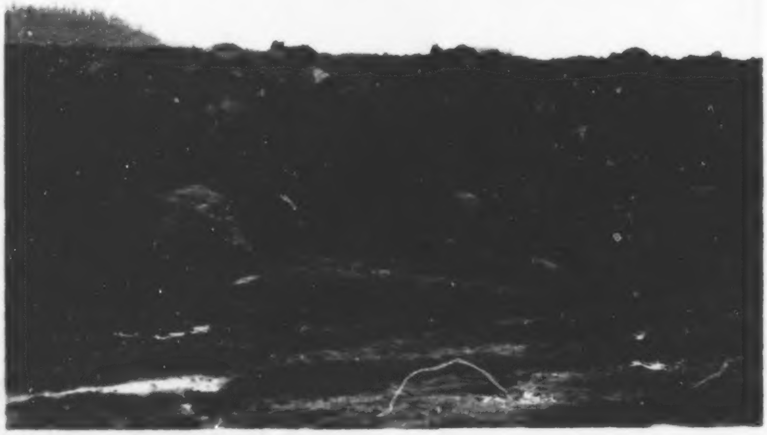


Figure 3. An overly thick lift of Poz-O-Tec containing "Boulders".

from the Title C hazardous waste regulations pending the results of an EPA study. In addition, the May 19, 1980 RCRA regulations modified the criteria for classification of hazardous wastes so that the leachate from the extraction test can be up to 100 times the Primary Drinking Water Standard. Very few coal combustion wastes would fail this criteria.

As a non-hazardous waste FGD sludge and ash are regulated under Title D of RCRA. The only potential area of concern in this writer's opinion is the permeability requirement for liners of 10^{-7} cm/sec. The Conesville field investigation has shown significantly higher permeabilities. The IUCS sludge fixation process still is considered generally more desirable from an environmental standpoint compared to direct ponding of unfixed sludge or thickening followed by ponding.

GYPSUM STACKING

One of the more promising disposal options to solve the problem of thixotropic FGD sludges is to provide for forced oxidation of the sludge to produce a waste gypsum. The thixotropic sulfite FGD sludges when oxidized to sulfate gypsum exhibit superior dewatering and handling characteristics. In addition, gypsum is a waste product of the phosphate fertilizer industry which has utilized stacking as a disposal method for over two decades.

EPRI, as a part of the Chiyoda Thoroughbred 121 FGD process demonstration at the Gulf Power Company Plant Scholz has funded geotechnical laboratory testing of the waste gypsum. The results of this field demonstration test indicate that FGD gypsum has the settling, dewatering, and structural characteristics similar to and, in some cases more favorable than phosphate gypsum, making stacking a viable option for disposal of FGD waste gypsum. The purpose of this paper will be to briefly summarize the EPRI gypsum stacking evaluation. A more detailed review of the evaluation is provided in the paper entitled "Evaluation of FGD Waste Disposal by Stacking," by T. S. Ingra, et al,[6] and the soon to be published EPRI final report (CS-1579, Vol. 3).[7] Copies of the Ingra paper are available at the FGD Symposium.

PROJECT PURPOSE AND OBJECTIVES

The objective of this EPRI project was to determine if the experiences of the phosphate fertilizer industry in stacking of gypsum could be applied to FGD by-product gypsum. To meet the objectives, two methods of evaluation were utilized to assess the stacking characteristics of FGD gypsum. Firstly, detailed geotechnical/civil engineering laboratory testing was conducted on Chiyoda Thoroughbred 121 from the Chiyoda pilot plant in Japan as well as the Scholz Power Plant of Gulf Power Company in Sneads, Florida to assess the physical and chemical properties, sedimentation, consolidation behavior, permeability characteristics, and shear strength properties relevant to stacking methods of waste disposal. During this phase of the investigation, test data was compared with similar data from gypsum produced by the phosphate fertilizer industry.

Secondly, a prototype FGD gypsum stacking was constructed and operated for a nine-month test period at Plant Scholz. The completed stack was approximately one-half acre (2023 m^2) and 12-feet (3.7 m) high. The effect of the addition of fly ash to FGD gypsum on its stacking characteristics was also investigated, because of the potential for simultaneous disposal of fly ash and gypsum.

CHARACTERISTICS OF FGD BY-PRODUCT GYPSUM

This paper highlights the detailed laboratory and field testing of the FGD gypsum performed as part of the stacking evaluation. Emphasis was placed on those chemical and civil engineering properties important to stacking methods of waste disposal.

Minerological Analysis

X-ray diffraction data from both the Japanese pilot plant CT-121 FGD gypsum and from samples from Plant Scholz indicated that gypsum ($Ca\ SO_4 \cdot 2\ H_2O$) was the only crystalline phase present. The X-ray diffraction trace was virtually identical to the one obtained from analytical reagent grade gypsum.

The morphology (crystal structure and form) of FGD gypsum crystals were evaluated by scouring electron photomicrographs. The gypsum crystals were found to be generally elongated with sharp, regular edges.

The crystals of Plant Scholz FGD gypsum varied in length from 0.05 to 0.25 min with an average of 0.13 mm, and varied in width from 0.04 to 0.06 mm. The length to width ratio averaged 2.7. The crystal rosette formation typical in many phosphate gypsums was not observed in either the pilot plant or Plant Scholz FGD gypsum.

Grain Size Distribution

Sieve and hydrometer analyses indicated that the FGD gypsum consists predominantly of non-plastic, poorly graded coarse silt size particles with a fines content of 100 percent (i.e., the percent by dry weight passing the U.S. No. 200 sieve.) The particle size distribution is shown in Fig. 4. The average particle diameter was found to be 0.06 mm.

Specific Gravity

The specific gravity of the FGD gypsum was found to vary from 2.27 to 2.44 with an average of 2.34 which agrees with the known specific gravity of gypsum of 2.33.

CIVIL ENGINEERING PROPERTIES

Sedimentation and Consolidation

During the study, the sedimentation-consolidation behavior of FGD gypsum was measured. Figure 5 shows the sedimentation-consolidation behavior of three FGD gypsum test samples. Two of these samples were allowed to settle in gypsum saturated water from the Plant Scholz gypsum stack and the remaining sample with distilled water. The range in initial void ratio and dry density were 0.88 to 0.91 and 12.25 kN/m^3 to 11.77 kN/m^3 (78.0 to 75.0 lb/ft^3) (71 to 73 percent solids) for the samples sedimented in gypsum-saturated water. Gypsum sedimented in the distilled water yielded a slightly higher initial void ratio of 1.04 and a corresponding lower dry density of 71.2 lb/ft^3 (69 percent solids).

The practical significance for the initial void ratio and dry density at low consolidation stresses is that the FGD gypsum will initially settle by gravity to a dry density of 11.77 kN/m^3 to 12.09 kN/m^3 (75 to 77 lb/ft^3) (71 to

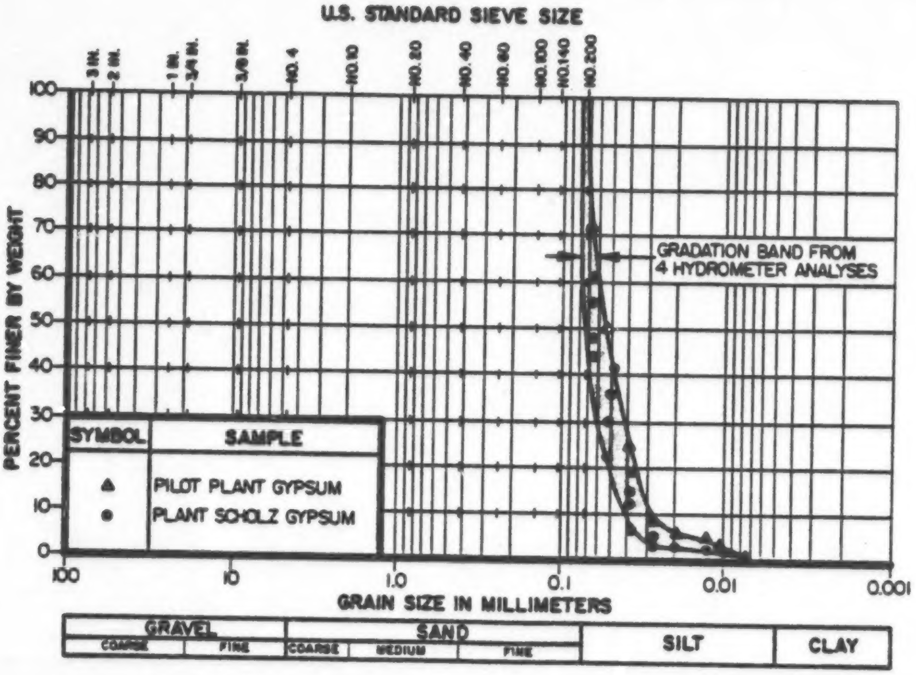
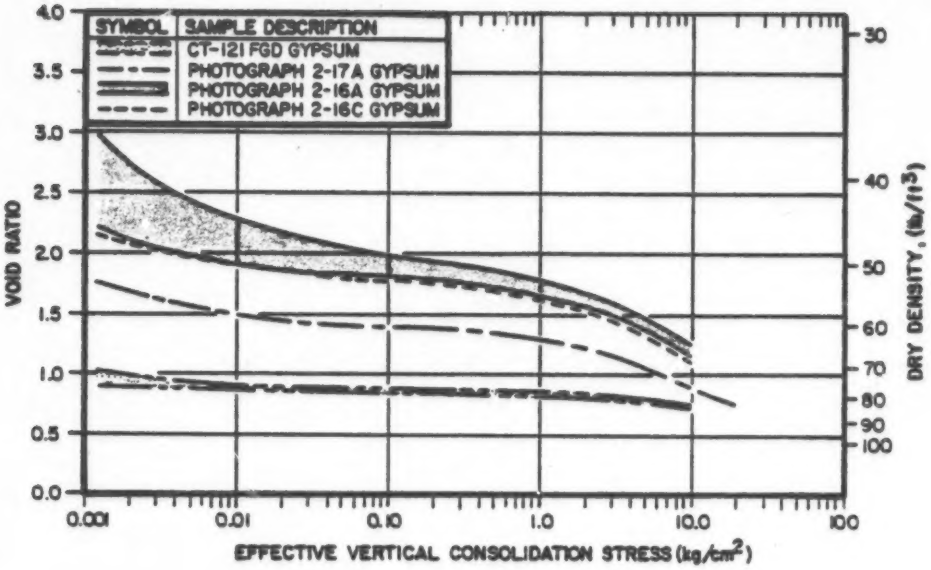


Figure 4. Grain Size Distribution of CT-121 FGD Gypsum
Source: EPRI Report No. CS-1579, Volume 3



1.0 lb/ft^3 = 0.157 kN/m^3
1.0 kg/cm^2 = 98.1 kPa

Figure 5. Sedimentation - Consolidation Behavior
Source: EPRI Report CS-1579, Volume 3.

72 percent solids) before consolidation under subsequent layers of sedimented gypsum.

Void ratio versus time consolidation curves for several stress levels for the Plant Scholz gypsum indicated that primary consolidation occurs quickly followed by secondary compression. The pH of the gypsum-saturated liquor was found to have little effect on the sedimentation and consolidation behavior of FGD gypsum.

Permeability

Permeability is an important material property from an engineering design standpoint as well as for its regulatory significance. Constant head permeability tests were performed on sedimentation column samples at low stress levels, and on cast and sedimented triaxial test specimens. Falling head permeability tests on undisturbed gypsum and in situ tests on sedimented gypsum were also run. As a result of variations in grain size distributions and crystal geometries, the coefficient of permeability for gypsum was found to vary by over one order of magnitude at equal void ratios or dry densities. Figure 6 compares FGD gypsum with the phosphate gypsums. The higher permeability of FGD gypsum is both a help and a hindrance. High permeability affects greater speed in dewatering and ease in handling which is important in constructing the containment dikes. The disadvantage is the increased quantity of leachate through the gypsum stack. The slightly higher permeability of the FGD gypsum versus the phosphate gypsum did not adversely effect stacking performance.

Shear Strength

Typically phosphate gypsum will have effective friction angles ranging from 45° to 50° measured by undrained triaxial compression test. The FGD gypsum was found to be similar to many phosphate gypsums and therefore acceptable for the stacking method of disposal.

Effect of Fly Ash Addition

Since some utilities may want to dispose of coal fly ash and gypsum together, the effect of fly ash addition on gypsum stacking characteristics was also

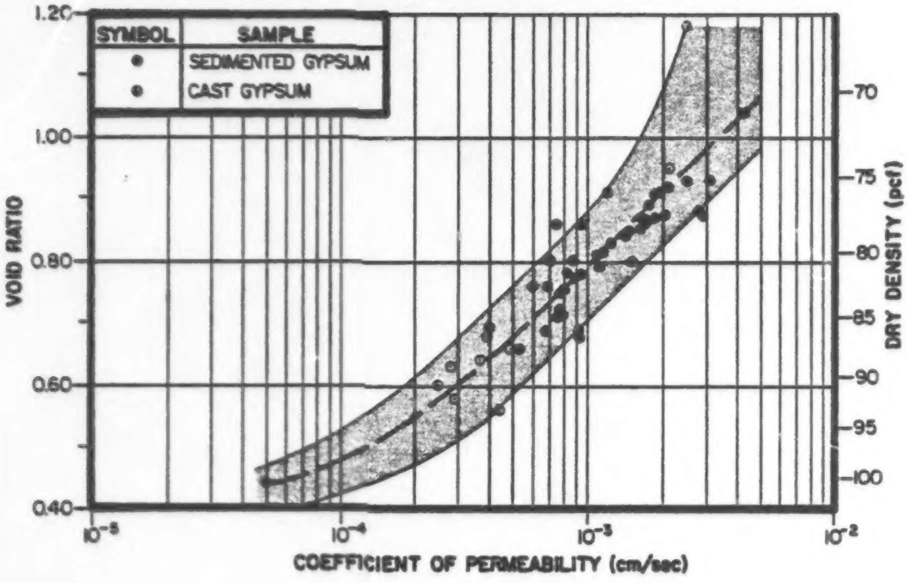


Figure 6. Void Ratio Versus Coefficient of Permeability
Source: EPRI Report CS-1515

studied. The results of laboratory testing on the fly ash/gypsum mixtures on the permeability, sedimentation-consolidation, and shear strength, characteristics all indicated reductions in the favorable stacking characteristics of FGD gypsum.

The strength characteristics of the fly ash gypsum mixtures appeared satisfactory for stacking provided the mixtures are properly drained and sedimented. On the other hand, the lower dry density, higher water content, and lower coefficient of permeability of the mixtures in comparison to the "pure" FGD gypsum will make excavation of sedimented material and construction of the perimeter dikes considerably more difficult.

STACKING DESIGN CONSIDERATIONS

Decades of practical experience in the phosphate gypsum industy indicates that the most economical method for disposal of FGD gypsum is by stacking using the upstream method of construction. In this method, an earthern starter dike is first constructed to create a sedimentation pond and stacking area. Gypsum is then pumped into the settling pond on slurried form (usually at between 15 to 20 percent solids) and allowed to drain. Process water is decanted and returned to the plant. When the gypsum sediments build up high enough, a dragline is used to excavate the gypsum to raise the perimeter dikes of the stack. The process of sedimentation, excavation, and raising of the perimeter dikes continues on a regular basis during the active life of the stack. Using this upstream method of construction, some gypsum stacks have reached heights exceeding 30 meters with slopes of 1.5 horizontal to 1.0 vertical.

Planning Aspects

The Scholz Plant disposal area and gypsum stack were proportioned for an estimated nine-month operation which would produce 5,500 to 6,500 tons (5.0×10^6 to 5.9×10^6 kg) and a final stack height between 7 and 8 meters. The geometry of the stack was governed by the minimum dimensions required for: (1) the safe operation of the dragline from the perimeter dike of the stack and (2) provisions for sufficient storage capacity within the center of the stack to allow clarification of the process water by sedimentation. Because of the relatively small amounts of gypsum produced in the test program compared to the magnitude of gypsum stacked in the phosphate fertilizer industry, the Scholz stack had to be small in area so that the stack could be raised enough to allow evaluation.

The planning of the gypsum stack and the process water return system included the following considerations: (1) estimate of gypsum production, (2) the planned life of the plant and the stack, (3) the planned life of in-flow and out-flow process water quantities and (4) estimates of water temperatures and climatic conditions. Once these processing constraints and climatic conditions were established, sizing of the gypsum stack and the process water system were made. The selected site plan and typical cross section of the disposal area and gypsum stack are shown in Figure 7. A liner was not used under the stacking area because the underlying soils were thought to be sufficiently impervious. In addition, the stack was only used for a short time before retirement and eventual removal.

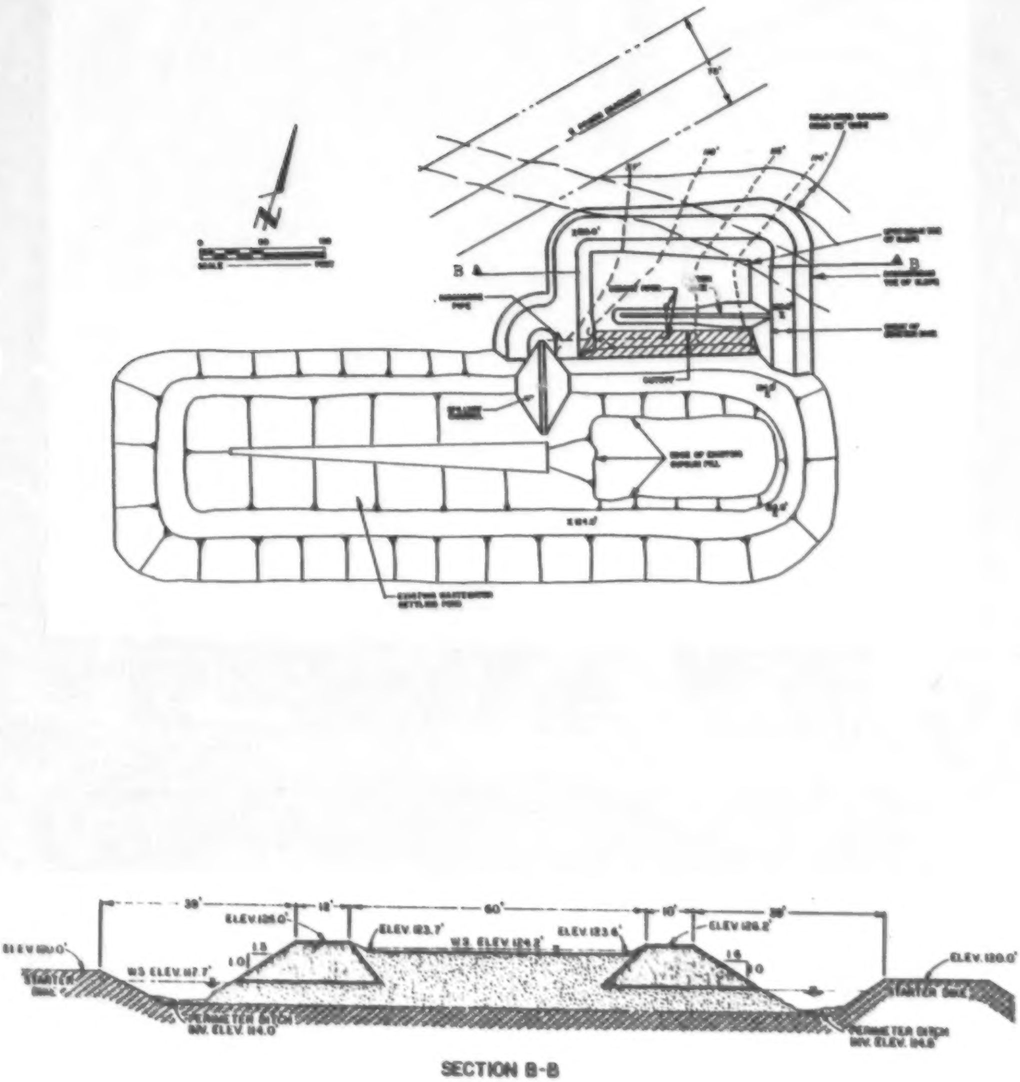
Stability

The Scholz FGD gypsum stack was raised a total of four times during the nine-month test program. As shown in the accompanying photographs, the sedimented gypsum can be walked upon at most locations within the diked containment area except near the slurry outlet.

There are two modes of failure which are normally considered in evaluating the stability of gypsum stacks[1], namely (1) a deep seated bearing capacity failure which is primarily controlled by the strength of the underlying foundation soils, and (2) seepage instability and progressive failure of the gypsum stack slopes which is controlled by the strength properties of the gypsum and the seepage pattern through the stack.

Process Water Return Systems

In the phosphate fertilizer industry, process water is normally pumped to the stacking area and is decanted from the settling waste gypsum and returned to the plant for reuse. The type and location of the decant system used to recycle the process water from the process to the plant can influence the quality and performance of the stacking operation.



B
B
30'
12'
60'
10'
30'
ELEV. 125.0'
ELEV. 123.7'
W.S. ELEV. 124.2'
ELEV. 123.6'
ELEV. 126.2'
ELEV. 120.0'
W.S. ELEV. 117.7'
1.5
1.0
STARTER DIKE
PERIMETER DITCH INV. ELEV. 114.0'
PERIMETER DITCH INV. ELEV. 114.0'
ELEV. 120.0'
STARTER DIKE
SECTION B-B

Figure 7. Plant Scholz Gypsum Stack Site Plan
and Typical Cross Section
Source: EPRI Report CS-1579, Volume 3

At Plant Scholz, a fixed vertical riser type decant structure was selected for the gypsum stack. This type of decant structure has the advantage of minimal maintenance requirements. The drawback to this type structure is that the spillway must be continuously raised as the sedimented gypsum rises. To avoid the disadvantage of the fixed vertical riser decant system a stage decant system can be used.

STACK CONSTRUCTION AND OPERATION

The construction and operation of a FGD waste gypsum stack using a dragline and the upstream method of construction was successfully demonstrated at Plant Scholz. During the testing at Plant Scholz efforts were made to construct and operate the stack as typically done in the phosphate fertilizer industry.

The gypsum stack was raised a total of four times by the upstream method of construction during the 9-month test program. The upstream method of construction is illustrated in Fig. 8. Photographs of the completed gypsum stack are shown in Fig 9. The height of the stack above the perimeter ditch averaged 3.7 meters (12 feet).

The structural integrity and environmental acceptability of the last Scholz gypsum stack was monitored throughout the 9-month test period from October 1978 to June 1977, and with additional field investigations continuing through June 1980.

The test program demonstrated that stacking of saturated FGD gypsum from beneath the water surface of an undrained pond using a dragline is practical provided the gypsum is placed on a dry surface and sufficient time is allowed for water to drain from the cast material before piling the gypsum more than a meter high. Dry gypsum located above the pond water surface was excavated and cast easily in stacks. As shown in Figure 10, the cast gypsum dikes of the stack were sufficiently stable and trafficable to allow the dragline to work upon the dikes without difficulty. Figure 11 shows the stack five months after the process shutdown, which when compared to earlier photographs shows no significant change in appearance of the north wall after several months of weathering and aging.

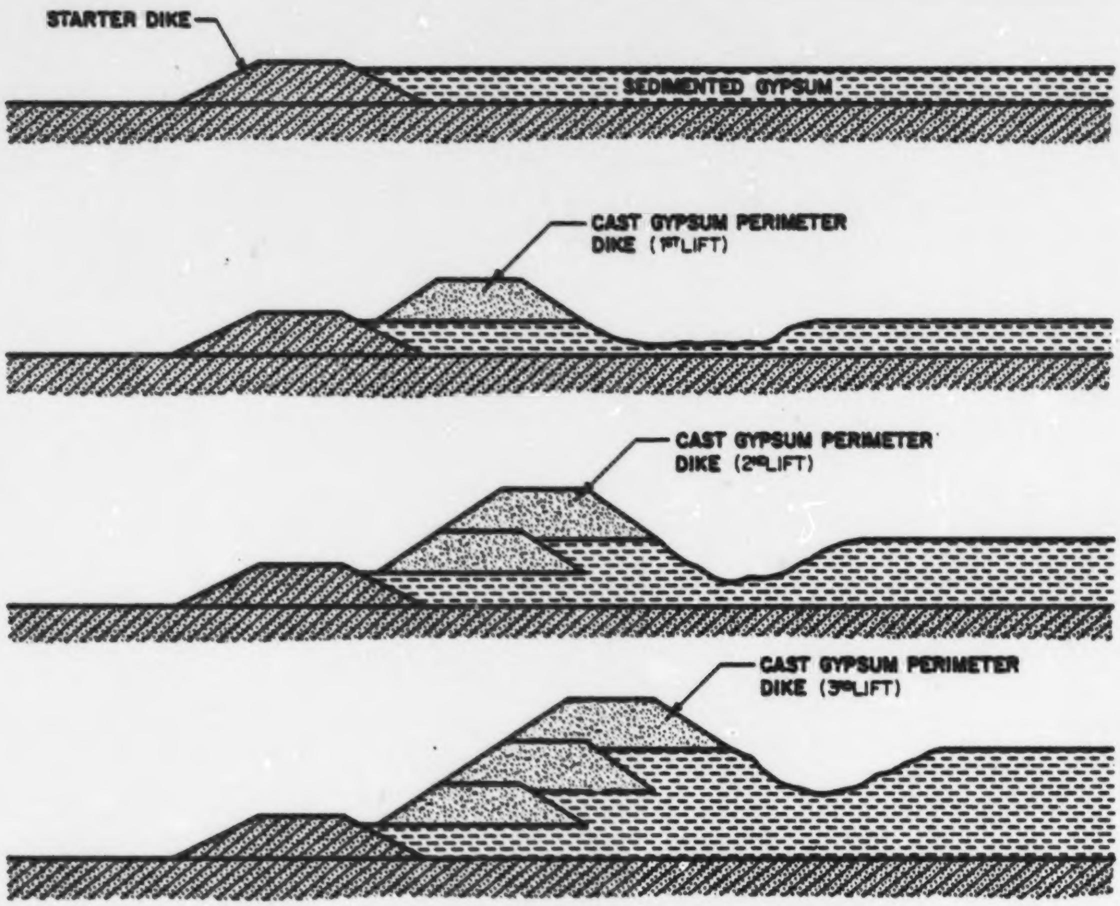
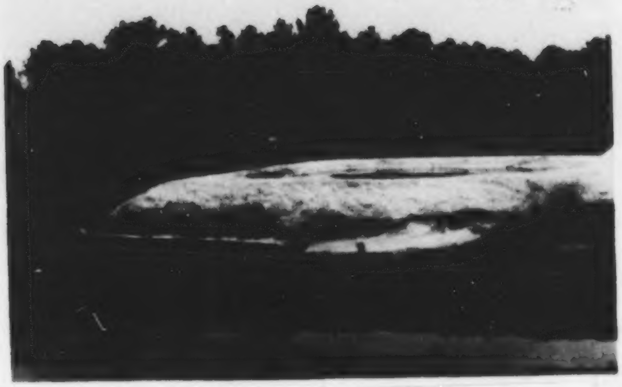
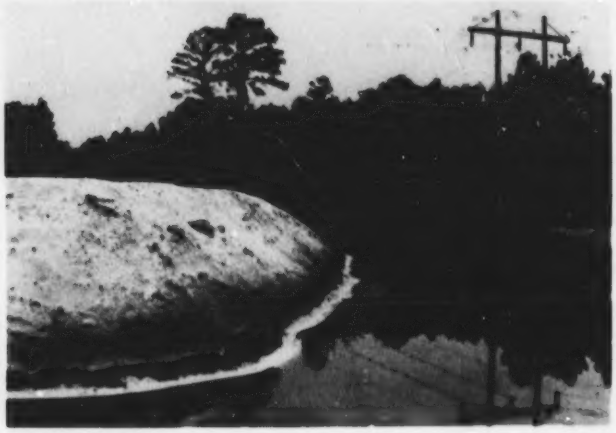

STARTER DIKE
SEDIMENTED GYPSUM
CAST GYPSUM PERIMETER DIKE (1ST LIFT)
CAST GYPSUM PERIMETER DIKE (2ND LIFT)
CAST GYPSUM PERIMETER DIKE (3RD LIFT)

Figure 8. Illustration of the Upstream Method of Construction
Source: EPRI Report CS-1515



(A) Overall View, July 1979



(B) North Wall, July 1979

Figure 9. Photographs of the Completed Gypsum Stack
Source: EPRI Report CS-1515

Figure 10. Dragline Operation from Crest of Cast Gypsum Dike
Source: EPRI Report CS-1515

Figure 11. North Wall of Stack 5 months after Process Shutdown
Source: EPRI Report CS-1515

The fly ash-gypsum mixtures did not stack as well as the gypsum alone. The poor drainage characteristics of the mixture and its higher water content, generally produced a much flatter cast slope than with the gypsum alone. Laboratory testing of the chemical and engineering properties of the FGD gypsum-ash mixtures also indicated a general reduction in the favorable stacking characteristics of FGD gypsum.

As was the case with the phosphate gypsum stacks, the cast FGD gypsum dikes and slopes developed a hard drying crust. The crust protect the slopes from erosion during rainfall and eliminate any dusting problem.

Groundwater Monitoring

Since the process water from FGD scrubbers can be of nearly any pH, and saturated gypsum and contains high concentrations of calcium and sulfate, there is the potential for groundwater contamination. For this reason, observation wells and piezometers were installed around the gypsum disposal area to monitor changes in groundwater quality during the test period. No "impermeable" liner was installed since the gypsum will be removed for sale after the project is completed. The observation wells and piezometers were installed close to the stack to determine changes in water quality rapidly. Water quality samples were obtained monthly during the active life of the stack and at six-month intervals afterwards.

The process water reaching the disposal area was also monitored on a monthly basis during the test period. The results from these analyses are summarized in Table 4.

TABLE 4

PROCESS WATER CHEMICAL COMPOSITION

Parameter	Average Test Period Value
pH	7.4
Ca^{++}	740 mg/l
Mg^{++}	780 mg/l
Na^{+}	90 mg/l
Cl^{-}	890 mg/l
$SO_4^{=}$	3050 mg/l
NO_3^{-}	530 mg/l
TDS	8900 mg/l

The water quality data indicated that leachate had entered an aquifer immediately below the stack as evidenced by consistent increases in all monitored parameters in wells below the stack. The FGD scrubber process waters contain concentrations of sulfate, calcium, chloride, nitrate, magnesium and sodium several orders of magnitude greater than secondary drinking water standards. Trace elements such as arsenic, chromium and selenium are also present within the process water at concentrations above the primary drinking water standards and could pose a potential problem. Therefore for stacking of FGD gypsum to be environmentally acceptable over the long-term operation of a full-scale stack, the seepage of leachate will need to be controlled or prevented. The recently published EPRI FGD Sludge Disposal Manual, 2nd Edition, (CS-1515, September 1980)[1] contains examples and illustrates of methods that can be used to effectively control leachate migration from gypsum stacks.

GYPSUM UTILIZATION POTENTIAL

Although evaluation of the utilization potential of FGD gypsum was not part of the stacking demonstration at Plant Scholz, it is relevant since the gypsum stacks would become storage areas rather than disposal sites. The utilization potential of by-product FGD gypsum has been evaluated by Chiyoda International Corporation as part of the overall CT-121 process evaluation. Uses were found in the manufacture of wallboard and Portland cement, and as an agricultural soil amendment.

Two full-scale production runs were made by U.S. Gypsum Company, Jacksonville, Florida using approximately 100 tons of CT-121 FGD gypsum each. The results indicate that this FGD gypsum could be used to manufacture wallboard to an equivalent quality of wallboard manufactured with natural gypsum.

Laboratory tests performed by Flinkote Company, Calaveras Cement Division confirmed that FGD gypsum is as acceptable as natural gypsum as a retarder in Portland cement.

Studies conducted by the University of Florida Agricultural Research and Education Center in Quincy, Florida indicates that FGD gypsum is a viable source of calcium and sulfur for peanut and soybean crops.

SUMMARY AND CONCLUSIONS

This paper has discussed results of field testing of two distinct methods of disposal of FGD by-products. The Conesville fixed FGD sludge landfill evaluation on the first full-scale use of the fixation process has uncovered a number of operational problems which have impacted the physical properties of the sludge/fly ash/lime mixture. Gypsum stacking on the other hand has been widely used in the phosphate industry. Presently there is approximately 300 million tons of waste gypsum which have been disposed of in Florida alone. The Scholz stacking demonstration has shown that the technology developed in the phosphate industry can be applied to utility FGD gypsum.

From a resource recovery standpoint the use of forced oxidation to produce a gypsum sludge and separate disposal of coal ash simplifies the utilization of these by-products. On the other hand, fixating sludge and with lime and fly ash eliminates any future re-use of the mixture.

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POTENTIAL EFFECTS ON GROUNDWATER OF FLY ASH AND FGD WASTE DISPOSAL IN LIGNITE SURFACE MINE PITS IN NORTH DAKOTA

by

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ABSTRACT

Increased reliance upon coal-burning power plants is resulting in the generation of large quantities of waste products. Fly ash and flue-gas-desulfurization (FGD) waste constitute the two major by-products of coal-burning power plants in the U.S. and at mine-mouth power stations a common method of disposal of these wastes operations is by emplacement in surface mine pits.

In this study, initially funded by EPA and presently by DOE, the potential impacts of surface mine pit disposal of fly ash and FGD waste at the Center Mine near Center, North Dakota, are being evaluated. The FGD waste at the Center site is generated by using the highly-alkaline fly ash as the SO_2 sorbent. The research involves field studies, laboratory studies, and computational geochemical studies to determine the potential for FGD waste and fly ash to affect groundwater quality.

FGD waste from North Dakota lignite, when placed in contact with water in various types of laboratory experiments, produces leach fluids with high SO_4, Mg and Na concentrations and pH values in the range of 7.0 to 8.5. Toxic metal and non-metal concentrations are generally not significantly in excess of drinking water standards. The concentrations of SO_4 and Ca are limited by the solubility of gypsum in conjunction with the degree of ion pairing between the major cations and sulfate. Highest total dissolved solids occur when Na and Mg are present as soluble sulfate salts.

In contrast, North Dakota lignite fly ash produces leach water that is very toxic because of extremely high pH and very high concentrations of As, Se and Mo. Analyses of groundwater samples from piezometers in unmined areas, in spoil with no buried waste and in spoil with waste have provided confirmation of the expectations of groundwater quality degradation based

on the laboratory experiments. Piezometers in and very near zones of buried fly ash yielded severely contaminated waters with high-pH and high concentrations of As, Se and Mo, whereas water from piezometers in spoil with FGD waste contained high SO_4 and cation concentrations but showed no significant quality deterioration due to As, Se, Mo or other toxic metals. Groundwater influenced by FGD waste is similar to groundwater in some areas of spoil without waste, where SO_4 and total dissolved solids concentrations are high. At present the areal extent of groundwater contaminated by FGD waste or fly ash at the Center site is small, but will increase gradually as a result of groundwater flow and as a result of continued waste disposal.

Environmentally safe disposal of fly ash and FGD waste in surface mine pits will require careful site selection and selective placement of wastes and the various overburden materials.

INTRODUCTION

Increased use of coal for production of electricity is resulting in the generation of large volumes of solid wastes, mainly in the form of fly ash as a residue from coal burning and flue-gas-desulfurization (FGD) waste as a residue from control systems for atmospheric emissions. Placement of the wastes in mine pits prior to backfilling the pits with overburden is a common method for disposal of these wastes at power plants located adjacent to surface coal mines. Although this waste disposal practice has existed for many years, little is known about the effects of fly ash and FGD waste in mine pits upon the quality of groundwater in and adjacent to the disposal areas.

This paper presents a summary of the results obtained to date from an on-going investigation of the effects of these wastes on groundwater quality at the Center Mine, operated by Baukol-Noonan, Inc., near the town of Center in western North Dakota (Figure 1). The waste is generated by the Square Butte Electric Cooperative Milton R. Young Station. This is a mine mouth power plant that fires low sulfur lignite coal from the Center Mine. Unit #2, a 450 MW cyclone-fired boiler, is fitted with an SO_2 scrubber which utilizes the high alkalinity of the plant's fly ash as the SO_2 sorbent. It is anticipated that this type of scrubber system will see wide application in the future. Thus, this site offers an excellent opportunity to evaluate the effects of the resultant waste products on groundwater quality.

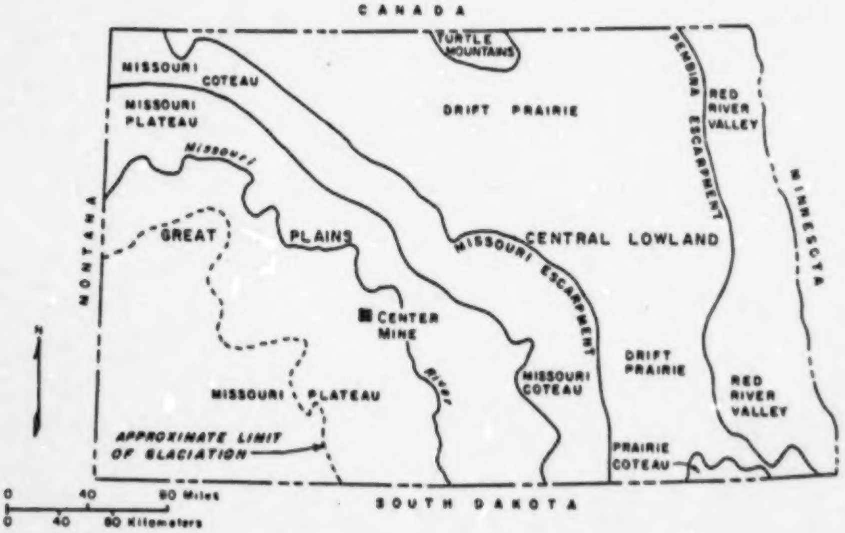
CANADA
TURTLE MOUNTAINS
MISSOURI COTEAU
MISSOURI PLATEAU
Missouri
DRIFT PRAIRIE
PEMBINA ESCARPMENT
RED RIVER VALLEY
MINNESOTA
MONTANA
GREAT PLAINS
MISSOURI ESCARPMENT
CENTRAL LOWLAND
CENTER MINE
River
N
MISSOURI COTEAU
DRIFT PRAIRIE
RED RIVER VALLEY
MISSOURI PLATEAU
APPROXIMATE LIMIT OF GLACIATION
PRAIRIE COTEAU
SOUTH DAKOTA
0 40 80 Miles
0 40 80 Kilometers

Figure 1. Location of Center, North Dakota study area.

The investigation was initially funded by EPA and is presently funded by DOE. Although the work pertains specifically to the wastes and their effects on groundwater at the Center mine, the results have relevance to the problems of fly ash and FGD waste disposal at other mines in the lignite and sub-bituminous coal region of the Northern Great Plains.

The investigation is comprised of laboratory experiments, field monitoring of groundwater in and near areas where fly ash and FGD waste have been buried, and hydrogeochemical computations to determine the influence of solubility constraints on the chemical evolution of groundwater affected by the wastes.

In the early part of the study, emphasis was placed only on determining the chemical composition of the FGD waste produced at the Center site and on determining the capability of this waste to cause degradation of groundwater quality. Later it became apparent that a considerable quantity of fly ash was being produced as a result of periods of inoperation of the flue-gas scrubber system. The fly ash, like the FGD waste, is disposed of by burial in the mine pits prior to backfilling with spoil. Although power plants equipped with fly ash scrubber systems will utilize much of the available fly ash, considerable quantities will remain requiring disposal. In addition, even with improved engineering, it is expected that periods of scrubber-system inoperation will always occur to some extent and thus fly ash is expected to continue to be a significant waste product. Thus, the influence of fly ash on groundwater became the second major aspect of this study. Most power plants in the United States have scrubber systems that use lime or limestone rather than fly ash as the SO_2 sorbent. Therefore these plants yield large quantities of fly ash as well as FGD waste products.

The hydrogeochemical properties of fly ash and fly-ash-type FGD waste are very different and these two waste forms can have quite different effects on groundwater quality. One of the purposes of this investigation is to determine if conversion of fly ash to FGD waste offers significant potential to reduce adverse effects on groundwater. Another reason for interest in fly ash is that large volumes of this product have already been disposed of by in-pit burial at older power plants where scrubber systems have not been installed or were not installed until recently.

DESCRIPTION OF THE FIELD SITE

The Center study area (Figure 2) is located on an upland divide bounded on the north and northeast by Square Butte Creek Valley and on the south by Hagel Creek Valley. Figure 2 shows the field study area, waste disposal site (active mining area), location of groundwater instrumentation in unmined areas adjacent to the disposal sites, and the location of the hydrostratigraphic cross section shown in Figure 3.

The climate in the Center area is semi-arid. The topography of the area is rolling. Glacial sediment (till) veneers much of the study area. The Tertiary coal-bearing strata consists of sand, silty sand, and lignite aquifers separated by silty and clayey materials (Figure 3). Glaciofluvial sand and gravel are present in the Square Butte Creek Valley.

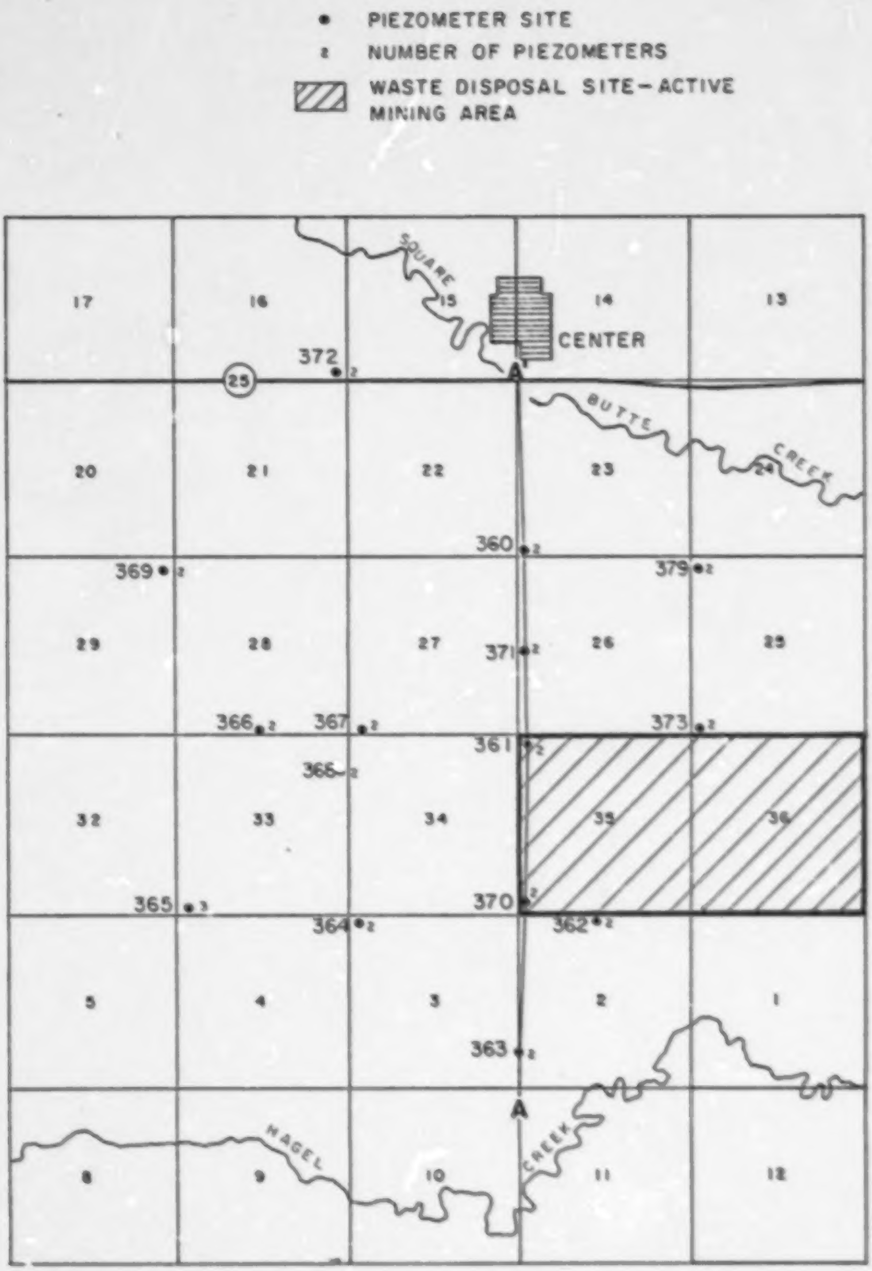
● PIEZOMETER SITE
2 NUMBER OF PIEZOMETERS
▨ WASTE DISPOSAL SITE—ACTIVE MINING AREA
17
16
15
14
13
SQUARE
CENTER
372
25
A'
BUTTE CREEK
20
21
22
23
24
360
369
379
29
28
27
371
26
25
366
367
373
361
368
32
33
34
35
36
365
364
370
362
5
4
3
2
1
363
A
HAGEL CREEK
8
9
10
11
12

Figure 2. Location of detailed study site, groundwater instrumentation in areas adjacent to study site, and location of cross section A-A' (Figure 3).

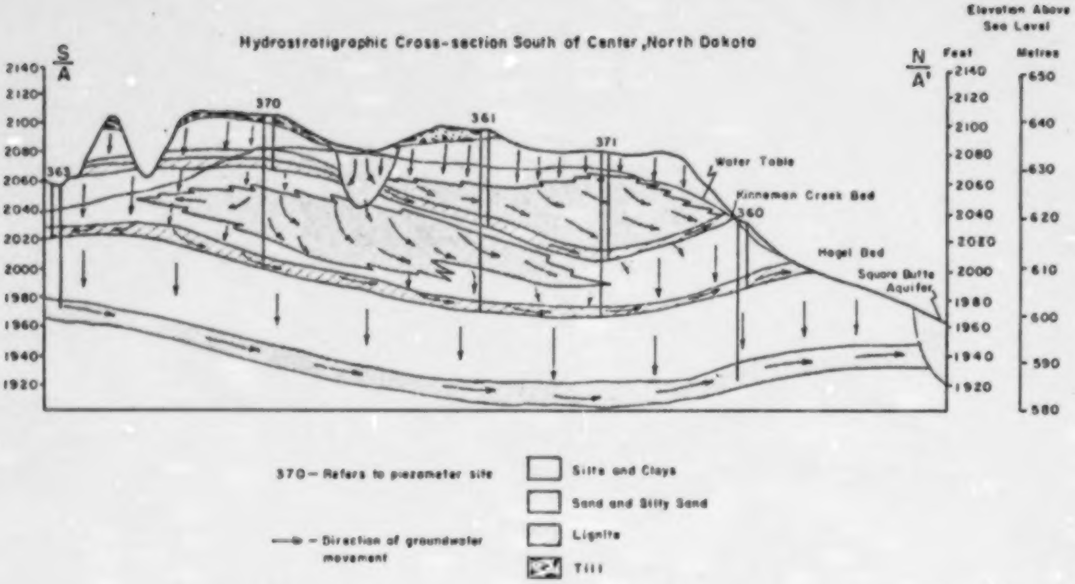
Hydrostratigraphic Cross-section South of Center, North Dakota
S
A
N
A'
Elevation Above Sea Level
Feet
Metres
2140
2120
2100
2080
2060
2040
2020
2000
1980
1960
1940
1920
650
640
630
620
610
600
590
580
363
370
361
371
360
Water Table
Kinneman Creek Bed
Hagel Bed
Square Butte Aquifer
370 — Refers to piezometer site
— Direction of groundwater movement
Silts and Clays
Sand and Silty Sand
Lignite
Till

Figure 3. Hydrostratigraphic cross section of the Center area. See Figure 2 for location of cross section.

The shallow stratigraphic units of hydrologic interest in unmined areas include two lignites; the Kinneman Creek Bed and the Hagel Bed, and a clayey sand unit which underlies the Hagel Bed (Figure 3). The Kinneman Creek and Hagel lignite beds are both mined at the present time. Much of the interval between the lignites consists of sandy materials as does part of the body of materials overlying the Kinneman Creek Bed.

The hydraulic interaction between the various stratigraphic units is shown in Figure 3. As is typical of most proposed and active surface mining areas in the Northern Great Plains, the Center site generally lies within an area of groundwater recharge. Therefore, waste products emplaced in the near-surface portion of the landscape lie in the path of infiltrating water and can potentially have a severe impact on groundwater quality.

Nearly all recharge to the groundwater system occurs during spring runoff and occasionally during periods of relatively heavy precipitation in the fall. Recharge occurs, to some degree, over most of the landscape but is concentrated in restricted positions. These include areas of standing surface water and ephemeral stream bottoms. In most other landscape positions, the majority of the precipitation is lost through either runoff or the process of evapotranspiration (Moran et al., 1978,[1] Groenewold et al., 1979[2]).

In general, the movement of groundwater in the fine-textured materials (aquitards) is vertical and downward. Movement of groundwater in the sand and lignite aquifers is essentially horizontal or lateral (Figure 3). Very slow downward movement of groundwater through the various aquitards supplies water to the aquifers. The aquifers, in turn, lose water by downward seepage to underlying aquitards and transmit water laterally to discharge areas along slopes, or, as in the case of the clayey sand unit, to glaciofluvial valley-fill materials (Figure 3).

The hydrologic units of major interest in the waste disposal area include the spoil materials and unmined units below the base of the spoils. FGD waste disposal was initially intended for two distinct positions within the mined area. These positions were the base of the mine pit and the V-notch between spoil ridges (Figure 4). However, due to numerous breakdowns in the scrubber system during the initial part of the study, nearly all of the FGD waste generated was emplaced in V-notch areas. Fly ash disposal was totally restricted to V-notch positions.

Typically, these two landscape positions are significantly different from the standpoint of groundwater considerations. The base of the mined area is relatively permeable and is generally below the postmining water table. The V-notch between spoil ridges is typically 10 to 20 metres above the base of the spoils and thus commonly above the position of the postmining water table. A major objective of the continuation of this study under DOE funds is a more detailed evaluation of the effect of FGD waste disposal in a saturated pit bottom area.

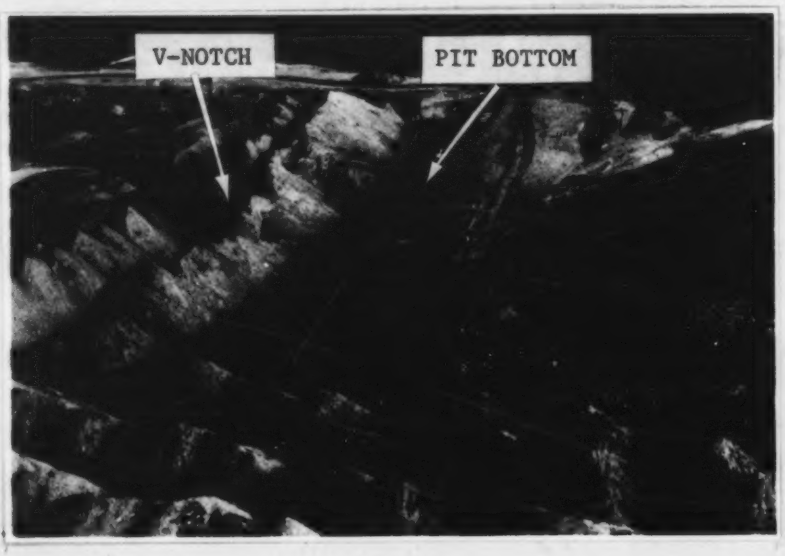

Figure 4. Photograph of Center Mine showing typical V-notch and pit-bottom disposal areas.

PHYSICAL PROPERTIES OF FLY ASH AND FGD WASTE

Each truckload of waste from the Milton R. Young Station consists of about two tonnes of slag and 18 to 22 tonnes of FGD waste. As received from the vacuum filter and subsequent to drying at 60°C, the solids content is 75%. The FGD waste is classified as silt (A-4) according to the AASH classification, and ML/OL by the Unified Soil Classification System. According to ASTM D-698, the maximum dry density is 90 pcf (1.44 gm/cm^3) and optimum moisture 22.5%. Permeability of field samples of FGD waste, as determined in the laboratory, is 6×10^{-6} cm/sec and under maximum density conditions is 2×10^{-6} cm/sec. A truckload of FGD waste and slag covers an area of approximately 14 feet x 14 feet (18.0 m^2) and is 5 feet (1.5 m) high in the middle. Initially the waste is very soft. However, after a few days of exposure, it will hold the weight of an adult person.

A truckload of FGD waste left exposed since January 1979 has shown very little change in surface appearance during the subsequent 16 months. It is possible to break off pieces of this material by hand and easily indent the surface. In the laboratory, a chunk of this exposed waste slaked in a few minutes when immersed in water. The dry density after 16 months exposure was 1.1 gm/cm^3 and the unconfined compressive strength was 116 psi or 8.1 Kg/cm^2. This is indicative of long term pozzolanic activity since the maximum unconfined compressive strength of moist cured cylinders compacted to ASTM D-698 (4 inch (10.16 cm) diameter cylinder, 4.5 inches (11.4 cm) deep, compacted with 5½ lb (2.49 kg) hammer, dropping 12 inches (30.48 cm), 25 blows per layer for 3 layers) only averaged 47.4 psi (3.3 kg/cm^2) after 9½ months.

During field sampling operations in October, 1979, it was found that driving 3 inch (7.62 cm) diameter thin-walled tubes through the buried FGD waste in a V-notch disposal area required much less effort than driving through the mine spoil below. In some parts of the V-notch as well as the pit bottom disposal areas, two or three layers of FGD waste have been emplaced. Since the rubber-tired dozers and trucks were prone to sink into the waste, even after several weeks of exposure, a 1 to 2-ft (30 to 60 cm) layer of mine spoil was required on top of each layer before driving on the waste.

The presence of about two tonnes of non-porous, coarse, glassy slag with each 18 to 22 tonnes of FGD waste causes porous channels and pockets for easy movement of water through the waste. The FGD waste also commonly shows extensive cracking after short periods of exposure in disposal areas. Where the sludge has been deposited only one truckload deep, numerous closed depressions between adjacent truckloads act as catchment basins and allow for the accumulation of water during heavy precipitation.

When the scrubber is not in operation the fly ash is diverted into a 2½-ft (76.2 cm) diameter x 8-ft (2.4 m) long rotary drum mixer inclined at 15°C. A shaft with blades attached uniformly mixes in about 15% water prior to dumping into the same trucks used for hauling FGD waste. In contrast to the FGD waste, the rubber-tired dozers are able to drive on the fly ash almost immediately without sinking in. After a few weeks of exposure, it is possible to break off pieces of the fly ash by hand and indent the surface, similar to the FGD waste that had been exposed for 16 months. However, even though the compressive strength of the fly ash is only 39 psi (2.7 kg/cm^2), there was only slight slaking when specimens were immersed in water.

In order to prevent slaking, the fly ash must have some degree of compaction, with at least 15% water, and be allowed to cure for about a week. In the laboratory, a series of compaction tests were performed at different compactive efforts and varying moisture contents. Even when compacted at optimum moisture content of about 15%, according to ASTM D-698, the test specimen that was immersed in water within an hour after compaction was completely slaked within a few minutes. However, similar specimens that were cured in sealed containers at room temperature for over a week and then immersed in water, experienced no slaking whatsoever. In addition, unusually high compressive strengths in the order of 2759 psi (193 kg/cm^2), were attained with this short curing period. Other specimens compacted at lower compactive efforts produced much lower compressive strengths. Specimens compacted to approximately the same density as the field sample (1.4 gm/cm^3) had compressive strength equal to 136 psi (9.6 kg/cm^2), and showed no signs of slaking.

Laboratory specimens of fly ash compacted to ASTM D-698 have permeabilities in the order of 2×10^{-7} cm/sec. However, from observations in the field it is evident that the field compacted fly ash is much more pervious and has numerous channels, pockets, and cracks. In contrast to the FGD waste, the fly ash displays very little slaking, provided the compacted fly ash-water mass has had several days of curing previous to being subjected to precipitation or infiltrating groundwater. If dry fly ash is simply dumped into a container or pool of water it will not set up or harden while immersed.

WASTE COMPOSITION

Fly Ash

Fly ash is the residue that remains after coal is burned for steam generation. It is a direct result of the occurrence of non-combustible inorganic impurities in the coal. The elemental composition of fly ash produced at the Milton R. Young power plant at the Center mine and, for comparison, the composition of fly ash produced by several other power plants in the Northern Great Plains region are shown in Table 1. The fly ash is comprised of numerous chemical compounds, the most abundant of which are silica (SiO_2) as glass, silica as quartz, $CaSO_4$ (anhydrite), Fe_2O_3 (hematite), Fe_3O_4, CaO, MgO, and C (carbon) The elemental composition of the fly ash produced at the Center site is generally similar to the compositions of fly ash generated by lignite-burning power plants elsewhere in the Great Plains region (Table 1). The composition differs significantly from that of fly ash produced from hard bituminous coal. The bituminous-coal fly ash has a higher silica (SiO_2) content, a higher iron content, and lower contents of calcium, magnesium and sulfur.

It should be noted that although elemental compositions of fly ash and FGD waste provide a useful means for comparison of these wastes, they do not indicate how the wastes will influence the chemical composition of water brought into contact with the wastes. In this regard we draw attention to the fact that mineralogical analyses of fly ash conducted in this investigation and those reported in the literature indicate that most of the sodium,

Table 1. COMPOSITION OF FLY ASH PRODUCED BY SEVERAL POWER PLANTS IN THE U.S.

Laboratory No.	Type of coal source for each fly ash	Chemical analysis (percent)								
		SiO_2	Al_2O_3	Fe_2O_3	S+A+F	Free CaO	Total CaO	MgO	SO_2	Alkalies as Na_2O
M 6498	Bituminous blend	46.1	19.0	18.6	83.7	2.2	8.2	1.3	1.6	0.72
M 6510	Lignite	37.1	11.8	7.3	56.3	0.5	21.8	5.6	2.6	4.23
M 6514	Lignite	37.2	15.5	5.6	58.3	0.2	24.3	11.3	0.9	0.07
M 6535	Subbituminous	28.8	20.0	4.1	53.0	2.1	32.0	6.4	3.8	0.68
M 6569	Subbituminous	51.8	27.2	2.0	81.0	1.9	10.7	2.1	0.7	0.85
M 6577	Lignite	31.1	17.1	7.9	56.1	0	25.3	8.1	3.3	1.35

Laboratory No.	Mineralogical analysis (percent) estimated[1]									
	CaO	Fe_2O_3	Fe_3O_4	Fe as Fe_2O_3	Quartz	MgO	Carbon	$CaSO_4$	Glass	Misc.
M 6498	4	4	4	8	4.0	1.5	4	2.5	68	10
M 6510	4	5	5	10	4.0	-	5	-	68	10
M 6514	5	2.5	2.5	5	4.0	2.5	2.5	2.5	68	10
M 6535	4	2.5	2.5	5	5.0	1.5	1.5	4.0	70	10
M 6569	4	2.5	2.5	5	2.5	4	4	2.5	70	10
M 6577	5	5	5	10	2.5	5	5	4.0	63	10

Laboratory No.	Composition of glass in ashes (percent)								
	CaO	Al_2O_3	Fe_2O_3	SiO_2	MgO	SO_3	Alkalies as Na_2O	Total	CaO/SiO_2
M 6498	4	19	11	42	0	0	0.72	77	0.10
M 6510	18	12	(-3)	33	6	-	4.23	73	0.55
M 6514	19	16	0.6	33	9	(-0.5)	0.07	78	0.58
M 6535	28	20	(-0.9)	24	5	1.4	0.68	78	1.17
M 6569	7	27	(-3)	49	(-2)	(-0.8)	0.86	84	0.14
M 6577	20	17	(-2)	29	3	0.9	1.35	71	0.70

[1] Estimated by X-ray diffraction and optical methods

M 6498	Chicago Fly Ash (Blend)
M 6510 (1973)	Basin Electric Power Corp., Stanton, North Dakota
M 6514 (1973)	Montana Dakota Utilities, Sidney, Montana
M 6535 (1974)	Public Service Co. of Colorado, Pueblo, Colorado
M 6569 (1974)	Navajo Generating Plant, Page, Arizona
M 6577 (1974)	Minnkota Power Co., Center, North Dakota

calcium, and magnesium in lignite fly ash is incorporated in glass and therefore is unavailable for dissolution in water. The forms of these elements that are significant with respect to the generation of total dissolved solids in groundwater are those occurring as oxides such as CaO and MgO and as sulfate salts such as $CaSO_4 \cdot 2H_2O$. Chloride salts occur in only very small concentrations in the lignite of the Northern Great Plains region and therefore these salts do not occur in the fly ash in significant amounts. The elemental composition of fly ash and FGD waste from the Milton R. Young plant is listed in Table 2. These data indicate that fly ash from the Milton R. Young plant contains measurable concentrations of numerous heavy metals, transition metals, and toxic non-metals such as arsenic. The fly ash also contains other constituents in significant concentrations that may cause degradation of groundwater quality.

FGD Waste

FGD waste is produced using fly ash to react with sulfur dioxide in the flue gas stream from the power-generation units. The fly ash, which is very alkaline, serves as a substitute for lime or limestone in the wet scrubber system for neutralization of sulfuric acid as sulfur is removed from the flue gas. Table 2 indicates that the composition of the FGD waste differs from that of the fly ash mainly by its high content of sulfate acquired from the flue gas. Another difference in the composition of the FGD waste and the fly ash, and one that is not evident from the elemental analyses, is the oxide content which is appreciable in the fly ash but which is very small in the FGD waste. Much of the calcium that exists as CaO in the fly ash occurs as anhydrite ($CaSO_4$) in the FGD waste.

FGD waste is composed primarily of quartz, glass, anhydrite, and iron oxides. Compounds of carbonate, fluoride, and chloride are present in only very small concentrations. Table 2 indicates that concentrations of heavy metals and toxic non-metals such as arsenic, selenium, and fluoride present in the FGD waste are similar to those present in the fly ash.

Calcium sulfate can occur in two forms, as gypsum, $CaSO_4 \cdot 2H_2O$, and as a nonhydrated salt, $CaSO_4$, known as anhydrite. At temperatures below about $40^{\circ}C$ at atmospheric pressure, gypsum is the thermodynamically stable compound. Anhydrite can exist, however, for long periods of time in a metastable state. Although the sulfate in the FGD waste is present primarily as anhydrite, a minor amount of gypsum probably also occurs.

Table 2 indicates that FGD waste produced at the Milton R. Young power plant at Center is somewhat variable in composition. This variability reflects the minor variations in the composition of the coal fed to the power plant. For reasons discussed below, this variability in FGD waste composition does not appear to be significant in terms of the influence of FGD waste on groundwater quality.

HYDROCHEMICAL LABORATORY INVESTIGATIONS

The influence of FGD waste on water chemistry was studied in the laboratory by analysis of or compilation of existing analyses of (i) supernatant

Table 2. CHEMICAL COMPOSITION OF FLY ASH AND FLY ASH FGD WASTE FROM THE MILTON R. YOUNG PLANT, CENTER, NORTH DAKOTA.

		mg/g									µg/g					
		Ca*	Mg*	Na*	K*	CO_3	Cl	SO_4	F	SO_3	Si*	Fe*	Cd	Hg	Se	As
	$\bar{x}$	122.1	35.2	28.8	12.5	0.5	0.01	93.4	0.2	0.2	258.3	63380	1.13	0.29	17.2	79.6
Fly Ash	s	14.7	2.0	10.9	0.3	0.1	0.00	12.7	0.4	0.04	10.6	2689	0.4	0.15	3.0	18.4
	n	5	5	5	5	5	5	5	5	5	5	5	5	5	5	5
	$\bar{x}$	104.8	19.0	15.0	13.6	3.2	0.01	195.6	0.3	0.1	166.3	50329	0.47	0.34	10.5	130.2
Fly Ash FGD Waste	s	24.8	3.0	4.7	4.3	8.1	0.00	57.5	0.06	0.1	39.8	6501	0.19	0.14	3.9	56.0
	n	17	17	17	17	17	17	17	17	17	17	17	17	17	17	17

* Calculated from % oxide

liquid from FGD waste storage vessels, (ii) samples of pore water extracted from FGD waste by squeezing, (iii) samples obtained from batch elutriation tests, and (iv) samples of effluent from column experiments. The first three types of analyses represent static (no flow) conditions, whereas the column experiments were designed to determine the changes in water chemistry that occur when waters of various initial compositions flow through samples of waste.

Process Liquors and Supernatant Solutions

Table 3 presents the results of an analysis of a sample of liquor collected from the absorber tower of the flue-gas desulfurization system at the Milton R. Young power plant and an analysis of a sample of supernatant liquid that occurred in a large vessel in which FGD waste from this power plant was transported from the field site to a laboratory. Although both samples are sulfate-salt solutions, the absorber liquid had a sulfate concentration of 65,500 mg/L whereas the supernatant liquid had a concentration of 5,130 mg/L. In each sample sodium was the most abundant cation, but it should be noted that the concentrations of calcium and magnesium were not determined.

The large difference in sulfate concentrations suggests that most of the sulfate in solution in the absorber tower precipitates from the system. For reasons described below, sulfate in water in contact with FGD waste that is discharged from the power plant does not rise to the levels measured in the absorber liquor. Providing that time for adjustment to chemical equilibrium is available, sulfate concentrations are limited by the solubility of gypsum in the presence of large concentrations of sodium and magnesium as well as calcium.

Pore Water From Squeeze Extraction

The FGD waste discharged from the power plant and transported to the mine pits contains pore water with dissolved solids. This water and the solid-phase composition of the waste represent the initial conditions of the water-and-FGD waste mass that is placed in the mine pits. To determine the chemical composition of this pore water, samples of FGD waste from field disposal sites at the Center Mine were collected on numerous occasions in 1978 and 1979. Pore water from the field samples of FGD waste was removed by squeezing, using a standard soil consolidation apparatus. For comparison, pore water from FGD waste used in leaching columns was also obtained by squeezing. The extraction method is described by Gullicks (1979[5]). The mean concentration of sulfate in the pore waters of the field samples was 10,763 mg/L with a range of 2,473 to 19,038 mg/L (Table 4). As is evident from Table 4, the dissolved constituents in the pore water of the field samples were at similar concentrations to those present in pore water squeezed from FGD waste used in leaching columns. The dominant cations were sodium, magnesium, and calcium. As was the case for the absorber tower liquor and the supernatant liquid, the pore waters were sulfate salt solutions. The pH of the pore water samples ranged from 7.5 to 8.9 with a mean of 8.23.

With the exception of sulfate, cadmium, fluoride, and lead, none of the mean values for the constituents included in the analyses exceeded the limits

Table 3. ANALYSES OF ABSORBER TOWER LIQUOR AND FGD SCRUBBER WASTE SUPERNATANT LIQUID FROM THE MILTON R. YOUNG PLANT.

ABSORBER TOWER LIQUOR[1]

Constituent	Concentration (mg/L)
Aluminum	329
Arsenic	0.42
Barium	0.16
Beryllium	0.015
Cadmium	0.27
Calcium	530
Cobalt	12.1
Chromium	0.16
Copper	0.4
Iron	808
Gallium	0.22
Lithium	2.5
Magnesium	9060
Manganese	63.4
Nickel	5.4
Potassium	1010
Silicon	293
Sodium	6760
Strontium	31.4
Uranium	2.7
Vanadium	0.34
Zinc	21.4
Chloride	200
Fluoride	0.55
Sulfate	65500
Sulfite	165

FGD SCRUBBER WASTE SUPERNATANT LIQUID[2]

Constituent	Concentration (ppm)
Bicarb. Alkalinity	0
Carb. Alkalinity	396
Chloride	74
Nitrate	8
Sulfate	5,130
T.D.S	18,576
Hardness	9,160
pH	8.9
Antimony	4.0
Arsenic	.0304
Barium	.50
Boron	290
Cadmium	.08
Calcium	--
Chromium (T)	<.05
Chromium (+6)	<.05
Cobalt	.60
Copper	.06
Cyanide	.05
Fluoride	15
Iron (T)	.16
Lead	.30
Magnesium	--
Manganese	.19
Mercury	.00125
Molybdenum	3.1
Nickel	.27
Potassium	200
Selenium	.088
Silver	.04
Sodium	1000
Zinc	.04

1. Data from Twin City Testing and Engineering Laboratory, Inc., 1976[3]

2. Data from Ness et al., 1978[4]

specified in the National Primary Drinking Water Standards. The recommended limit for sulfate in drinking water is 250 mg/L. In contrast, the sulfate concentrations in the pore-water samples are larger by a factor of 10 to 76. The mean value for lead is only slightly above the maximum permissible concentration specified in the drinking water standards. Fluoride concentrations in the pore waters ranged from 1 to 28 mg/L. In contrast, the maximum permissible limit specified for fluoride in drinking water is 2.4 mg/L in

Table 4. COMPARISON OF THE CONCENTRATIONS OF CHEMICAL CONSTITUENTS IN THE LEACHING COLUMN FGD WASTE POREWATER TO THE CONCENTRATIONS FOUND IN THE POREWATER OF FIELD SAMPLES.

	Porewater From 27 Samples of FGD Waste From Field Disposal Sites			Leaching Column FGD waste
Constituent	Range	Mean	Std. Deviation	Porewater
Arsenic, µg/L		Not Available		11.7
Cadmium, µg/L	1-156	18.6	38.7	547.0
Calcium, mg/L	298-530	423.2	53.0	395.0
Chromium, µg/L	11-63	32.1	14.8	26.9
Cobalt, µg/L	10-162	61.2	42.8	98.4
Lead, µg/L	17-219	88.0	55.8	58.0
Magnesium, mg/L	368-2903	1,633.2	802.1	2,310.
Mercury, µg/L		Not Available		0.7
Potassium, mg/L	7-354	44.3	105.5	332.0
Selenium, µg/L		Not Available		25.9
Sodium, mg/L	237-3320	1,623	1,054.7	3,200.
Total Alkalinity (as $CaCo_3$), mg/L	38-528	217.9	120.2	87.5
Chloride, mg/L	5-106	40.1	27.8	29.7
Fluoride, mg/L	1-28	13.7	7.1	5.0
Total Dissolved Solids, mg/L	4032-30,968*	17,064*	7,145*	28,273.
Sulfate, mg/L	2473-19,038	10,763	4,029	17,106.
pH	7.5-8.9*	8.23*	0.37*	8.10

* Based on 24 FGD waste porewater sample analyses

regions with a mean annual temperature of 12°C. The limit varies slightly based on the regional temperature which is taken as an index of drinking water comsumption. Cadmium concentrations in the pore waters from the field samples ranged from 1 to 156 µg/L. The mean cadmium concentration was 18.6 µg/L. The maximum permissible limit specified for cadmium is 10 µg/L. The cadmium concentration in pore water from the leaching column FGD waste was 547 µg/L.

Water from Elutriation Tests

A number of shaker-type elutriation tests were performed on FGD waste from the Milton R. Young plant. These tests were performed as part of a multi-laboratory test program designed to determine the reliability of three methods for hazardous waste evaluation. Results of the FGD waste elutriation tests are shown in Table 5. Of the constituents included in the analyses, none exceeded the limits specified in the National Primary Drinking Water Standards.

Similar shaker-type elutriation tests were performed on samples of fly ash from the Milton R. Young Plant. Results of these tests are shown in Table 6. Comparison of the data in Table 6 with those in Table 5 indicates significant differences in pH as well as arsenic and selenium concentrations. The pH from fly ash elutriation tests ranged as high as 11.46 compared to a high of 6.58 for FGD waste. Arsenic concentrations from fly ash elutriations range from 35.6 to 127 μg/L as compared to <2 μg/L for FGD waste. The maximum permissible limit specified to arsenic in drinking water is 50 μg/L. Selenium concentrations from fly ash elutriations range from 17.6 to 368 μg/L as compared to <2 μg/L for FGD waste. The maximum permissible limit specified for selenium in drinking water is 10 μg/L.

The concentrations of all other constituents analyzed were very similar for the fly ash and FGD waste elutriations. As discussed previously, the concentrations of all constituents in fly ash and FGD waste from the Milton R. Young plant, with the exception of SO_4, are very similar (Table 2). Thus, the relatively high concentrations of arsenic and selenium generated by elutriation of fly ash are particularly significant and suggest selective mobilization of these constituents from the fly ash.

Table 5. RESULTS OF A SHAKER-TYPE ELUTRIATION TEST OF FGD WASTE FROM THE SQUARE BUTTE ELECTRIC COOPERATIVE UNIT No. 2[1]

Parameter	Concentration in the Elutriate (Range of 3 Elutriation Methods Used
pH	4.99-6.58
Calcium mg/L	401-600
Chromium mg/L	.001-.016
Arsenic μg/L	<2
Barium mg/L	.225-.716
Cadmium μg/L	.5-2.7
Lead μg/L	<1.5-12.1
Mercury μg/L	<.3
Selenium μg/L	<2
Silver mg/L	.002-.024

[1] Performed by David J. Hassett, Chief Chemist, Engineering Experiment Station, University of North Dakota, April, 1979, as part of a multi-laboratory test program designed to test the effectiveness and reliability of three elutriation methods for hazardous waste evaluations.

Table 6. RESULTS OF A SHAKER TYPE ELUTRIATION TEST OF A LIGNITE FLY ASH FROM THE SQUARE BUTTE ELECTRIC COOPERATIVE UNIT No. 2[1]

Parameter	Concentration in the Elutriate (Range of 3 Elutriation Methods Used)
pH	6.96-11.46
Calcium mg/L	394-951
Chromium mg/L	.031-.160
Arsenic µg/L	34.6-127.0
Barium mg/L	.757-1.069
Cadmium µg/L	1.3-17.6
Lead µg/L	2.7-22.4
Mercury µg/L	<.3-0.4
Selenium µg/L	17.6-368.0
Silver mg/L	.007-.015

[1] Performed by David J. Hassett, Chief Chemist, Engineering Experiment Station, University of North Dakota, April, 1979, as part of a multi-laboratory test program designed to test the effectiveness and reliability of three elutriation methods for hazardous waste evaluations.

A simple experiment was devised to determine the potential mobility of arsenic from fly ash from the Milton R. Young plant. Ten samples of fly ash, collected from the plant on a monthly basis, were elutriated with distilled-deionized water. The results of this experiment appear in Table 7. The pH of the 10 resultant fluids ranges from 11.95 to 12.75. The arsenic concentrations ranged from 15 to 770 µg/L, suggesting that release of arsenic from fly ash from the Milton R. Young plant may constitute a significant concern.

Effluent from Leaching Columns

Leaching column experiments were conducted utilizing FGC waste from the Milton R. Young plant. Column experiments were not performed using fly ash. The flow-through leaching experiments were performed by packing 10 cm-long columns of FGD waste in translucent plexiglass tubes that were 1.5 cm in diameter. Each tube was fitted with machined plexiglass base plates with a rubber o-ring and an effluent discharge opening and a Teflon stopcork. An inert porous plastic filter pad was placed on the base plate to support the sample and retard passage of particulate matter from the column. The mass of FGD waste placed in each tube was compacted to achieve a wet density of approximately 1.7 g/cm^3, which was the wet density determined for FGD waste samples under field conditions.

The FGD waste used in the columns was compared to seventeen FGD waste samples from two disposal areas at the Center site and was found to have no

Table 7. ARSENIC CONCENTRATIONS AND pH OF FLUIDS PRODUCED BY ELUTRIATION OF FLY ASH SAMPLES.

Sample	pH	Arsenic (µg/L)
FA 1 (8/10/78)	12.75	33.0
FA 2 (9/11/78)	12.02	164.0
FA 3 (10/30/78)	12.06	217.0
FA 4 (11/15/78)	12.11	79.0
FA 5 (12/ 8/78)	12.28	15.0
FA 6 (1/31/79)	12.05	212.0
FA 7 (2/ 9/79)	11.95	770.0
FA 8 (3/ 8/79)	12.28	433.0
FA 9 (4/11/79)	12.21	70.0
FA 10 (5/ 1/79)	12.03	15.0

1. Each fly ash sample was weighed into a polycarbonate centrifuge tube, about 6.3 g of ash in each tube.
2. 25 ml of distilled deionized water were added to each tube.
3. The tubes were gently agitated with inversion for 24 hours ± 0.5 hours.
4. Contents of the tubes were centrifuged and the clear upper layer was analyzed.

significant differences in chemical composition. Fourteen columns were used, each with a different type of leaching fluid as follows: distilled-deionized water, groundwater from the Hagel lignite bed (the main lignite aquifer at the Center site), groundwater from the base of the spoil in one of the mined segments at Center, water from the effluent weir of the bottom-ash settling tank at the Milton R. Young plant, and ten waters obtained by elutriation of samples of five different types of overburden from the Center site with distilled-deionized water and groundwater from the Hagel lignite bed. Prior to passage of the various influent liquors through the columns, a filtered (0.45 µm) sample of each was analyzed for total dissolved solids, pH, potassium, sodium, magnesium, calcium, chloride, sulfate, alkalinity, fluoride, cobalt, cadmium, chromium, and lead.

The ten elutriation fluids were produced by stirring a mixture of one part overburden material with four parts water. After 16 to 20 hours of stirring, the mixture was allowed to settle and the liquid was removed by decantation. The decanted liquid was then added to fresh overburden material in the same proportions indicated above and was shaken in a wrist-action shaker until measurements of pH, electrical conductance, sodium, and total dissolved solids were found to indicate a stable condition. Before the fluids were used in the columns, they were passed through 2 µ filter paper.

The results of chemical analyses of the elutriation fluids and of the other fluids used in the column experiments are listed on Table 8. Of the

Table 8. ANALYSES OF INFLUENT LIQUORS FOR LEACHING COLUMN EXPERIMENTS

Analyses of the ten types of elutriation fluids produced for use as leaching column influent liquors.

Type of Leaching Column Influent Liquor:	Hagel Lignite Bed Groundwater Elutriated With					Distilled-Deionized Water Elutriated with				
	Oxid. Sand	Oxid. Glacial Till	Oxid. Brown Clay	Unox. Grey Clay	Unox. Silt	Oxid. Sand	Oxid. Glacial Till	Oxid. Brown Clay	Unox. Grey Clay	Unox. Silt
Elutriations:	2	3	3	2	2	2	4	3	2	2
Constituent										
Cadmium, µg/L	<1	<1	<1	<1	<1	<1	1.4	<1	<1	2.1
Calcium, mg/L	11.0	364.	212.	17.0	8.4	28.7	403.	191.6	13.2	4.36
Lead, µg/L	7.1	8.4	3.1	9.4	12.8	0.8	15.7	9.2	14.6	19.0
Magnesium, mg/L	33.2	260.9	163.4	16.4	13.4	12.0	271.5	126.8	5.59	4.17
Potassium, mg/L	7.4	11.2	20.8	22.8	18.0	1.4	9.2	14.0	11.2	11.2
Sodium, mg/L	248.	420.	264.	296.	304.	36.	240.	76.	156.	134.
Total Alkalinity, as ppm $CaCO_3$	227.4	175.5	222.3	376.6	399.8	43.7	38.6	57.1	197.6	153.3
Chloride, mg/L	8.9	21.1	9.9	7.6	8.9	3.2	16.2	3.9	2.1	3.4
Fluoride, mg/L	0.7	0.8	0.8	1.0	0.8	0.8	1.1	0.7	1.3	1.0
Total Dissolved Solids, mg/L	1016.	4135.	2599.	1078.	1121.	186.	4248.	1943.	381.	563.
Sulfate, mg/L	408.	2417.	1070.	331.	401.	172.	2645.	891.	146.	121.4
pH	8.7	8.5	8.6	8.8	8.7	9.0	8.5	8.3	8.7	8.7

Table 8 (continued). ANALYSES OF INFLUENT LIQUORS FOR LEACHING COLUMN EXPERIMENTS.

Analyses of leaching column influent liquors other than those produced by the elutriation of overburden materials

Constituent	Distilled Deionized Water	Hagel Lignite Bed Groundwater	Groundwater From the Base of the Spoil Piles	Water From the Effluent Weir of the Bottom Ash Settling Tank
Cadmium, μg/L	<1	<1	<1	<1
Calcium, mg/L	6.7	11.0	138.4	137.
Lead, μg/L	0.6	2.3	6.1	6.5
Magnesium, mg/L	0.5	6.8	150.1	66.2
Potassium, mg/L	0	4.6	20.8	26.4
Sodium, mg/L	0	384.	176.	360.
Total Alkalinity, as ppm $CaCO_3$	4.1	747.6	277.8	64.3
Chloride, mg/L	<0.1	3.8	5.9	9.3
Fluoride, mg/L	<0.1	1.3	<.1	0.8
Total Dissolved Solids, mg/L	24.	1,147.	1,947.	2,011.
Sulfate, mg/L	2.0	122.	1,056.	1,090.
pH	5.7	8.4	8.1	8.2

five types of overburden materials used to obtain elutriation fluids, two (the glacial till and the oxidized brown clay) produced solutions that contained high concentrations of total dissolved solids and sulfate. These high concentrations occurred for both types of elutriation solutions, the groundwater from the Hagel lignite bed and distilled-deionized water. These results suggest that some of the types of overburden in the Center area are capable of causing poor quality groundwater to occur in spoil where fly ash or FGD waste are absent. Field evidence in support of this conclusion is presented in a following section of this paper.

The other three types of overburden used to obtain the elutriation fluids yielded solutions with total dissolved solids between 1016 and 1121 mg/L when groundwater from the Hagel lignite bed was used for elutriation and values between 186 and 563 mg/L when distilled-deionized water was used (Table 8). The total dissolved solids of groundwater from the Hagel lignite bed used for elutriation was 1147 mg/L. Although the elutriation fluids had similar values of total dissolved solids as the Hagel lignite bed groundwater, they had more than twice the sulfate concentrations and half or less than half the alkalinity of the groundwater from the Hagel Bed.

All influent liquors were added to the columns until an overflow level was attained. The columns were filled daily to the overflow level, after recording the fluid level decline of the previous day. For most columns, effluent representing the first seven pore volumnes, the eleventh through seventeenth pore volumes, and the forty-fourth through fiftieth pore volumes were chemically analyzed. For some of the columns, experimental problems necessitated that other pore volume samples be analyzed.

The results of a typical leaching column experiment are shown in Figure 5. The column experiments indicated that the concentrations of sodium and sulfate increased to approximately the same high levels in the initial pore volumes of all the columns, regardless of influent chemistry. However, by 50 pore volumes, the sodium concentrations generally decreased to approximately the concentrations of the influent. The sulfate concentrations by 50 pore volumes were generally slightly higher than influent concentrations. The magnesium concentrations always increased to approximately the same levels in the first few pore volumes, regardless of influent chemistry. However, by 20 pore volumes the magnesium concentration typically decreased to approximately that of the influent. The calcium concentrations in all the columns attained approximately the same levels in the first few pore volumes, regardless of influent chemistry. The calcium concentration typically showed a slow decrease from 10 to 50 pore volumes. In contrast to the other ions, the calcium concentrations at 50 pore volumes were typically 2 to 10 times greater than the influent concentrations for calcium.

These data indicate that the FGD waste contains soluble compounds of magnesium, sodium, and sulfate. These soluble salts are generally flushed from the columns by 10 pore volumes. Calcium concentrations increase to a maximum from pore volumes 1 to 10. Thereafter, the effluent becomes saturated with respect to gypsum. Once the magnesium and sodium compounds are flushed, only gypsum dissolution contributes to the column effluent.

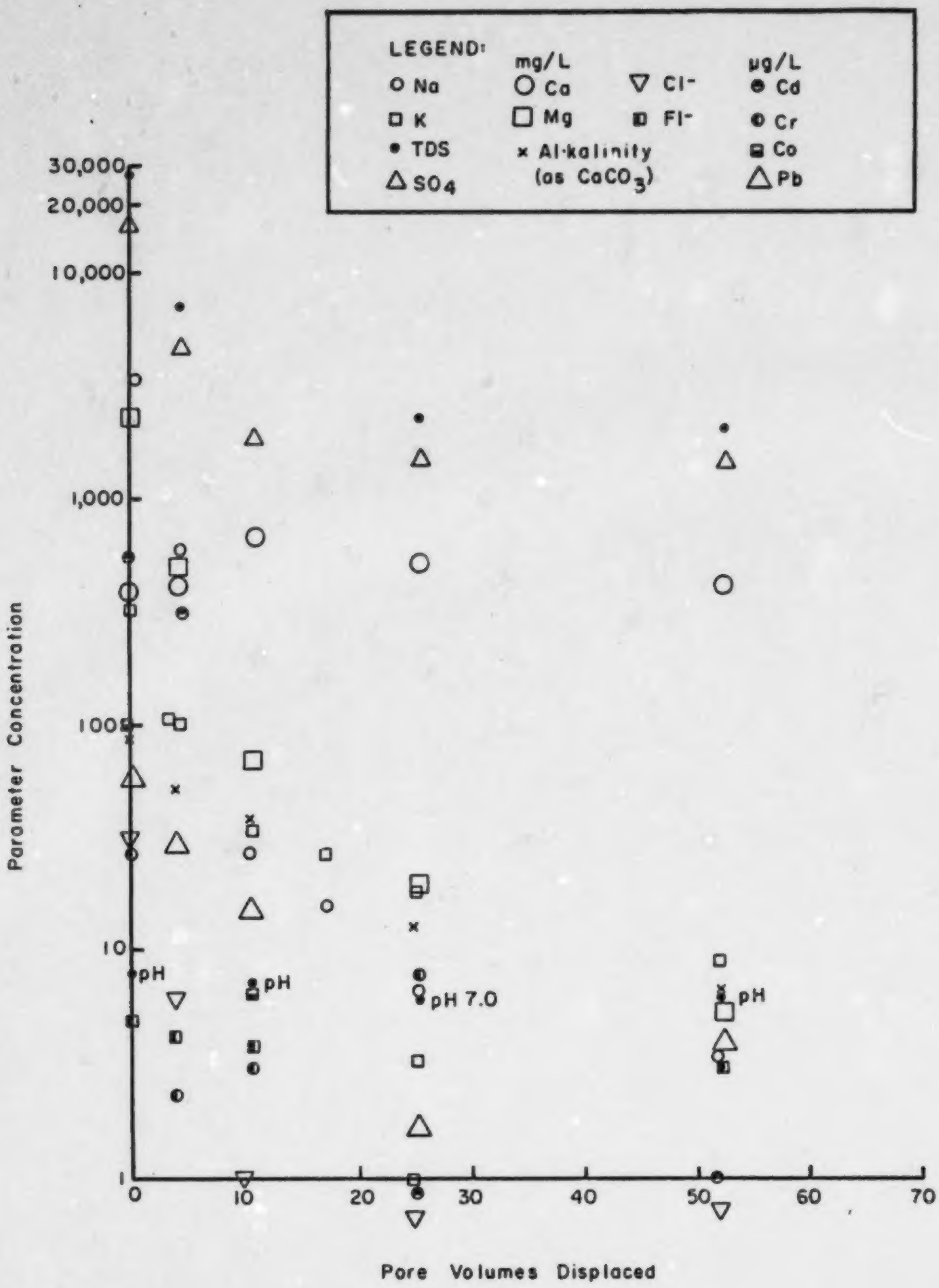

Figure 5. Analytical Parameters of column A011 versus average pore volume displacement.

HYDROGEOCHEMICAL PROCESSES

The results of a representative group of the chemical analyses of pore water squeezed from FGD waste, FGD waste column effluent, and contaminated and unaffected spoil groundwater were processed using a computer program known as WATEGM for equilibrium geochemical computations. With the analytical results as input, this program computes the distribution of chemical species and determines saturation indices for selected solid phases. The thermodynamic basis for a program similar to WATEGM is described by Truesdell and Jones (1974[6]). Documentation and explanation of WATEGM is currently being prepared as part of a PhD thesis by Carl D. Palmer, University of Waterloo, Waterloo, Ontario. Representative WATEGM results are listed in Table 9, which shows that a characteristic feature of water in contact with FGD waste and fly ash is that it is at or very near equilibrium with respect to gypsum ($CaSO_4 \cdot 2H_2O$). Although the solid-phase form of the calcium sulfate in the FGD waste and fly ash is anhydrite ($CaSO_4$), the equilibrium control on calcium sulfate in solution is gypsum, because gypsum is the thermodynamically stable phase at temperatures below about 40°C.

Although the waters are in equilibrium with respect to gypsum, there is considerable variability in the concentrations of sulfate and calcium in the various solutions (Figure 5). This variability is attributed to the presence of relatively large concentrations of sodium and/or magnesium. These cations form pairs or complexes with sulfate, such as $MgSO_4^o$, $NaSO_4^-$, and $Na_2SO_4^o$, which greatly increase the solubility of gypsum because a large percentage of the total dissolved sulfate exists in these solution species (Table 9). Sulfate is the main component of total dissolved solids of water in contact with FGD waste and fly ash. Because equilibrium sulfate concentrations can vary considerably as a result of differences in sodium and magnesium in the water, there is considerable variation in the total dissolved solids of waters in contact with FGD waste and fly ash.

Sulfate salts of sodium and magnesium, such as thenardite (Na_2SO_4), mirabilite ($Na_2SO_4 \cdot 10H_2O$), and epsomite ($MgSO_4 \cdot 7H_2O$) are extremely soluble in water. The concentrations of these two cations in pore water squeezed from FGD waste, FGD waste column effluent, and groundwater influenced by fly ash burial are much below their sulfate-salt solubility limits (Table 9). The concentrations are probably limited by the amounts of sodium and magnesium in soluble solid phases in the FGD waste or fly ash which, for reasons presented below, are much less than total amounts of these elements in the wastes. Figure 5 indicates that as water was passed through the columns of FGD waste, the concentrations of sodium and magnesium rapidly declined to levels much below those for the initial pore volumes. This is particularly the case for magnesium, which for many of the columns generally decreased to concentrations of less than 100 mg/L and in a few cases less than 20 mg/L from initial values generally greater than 2000 mg/L. Mass balance calculations for the column effluent indicate that although the concentrations are much lower than the initial values after 50 pore volumes were passed through the columns, less than 20% of the total mass of these elements within the FGD waste was removed. This indicates that most of these elements in the columns exists within relatively insoluble solid phases such as the glass component of the FGD waste. In this context, elemental analyses of FGD waste are of only limited value for use in

Table 9. REPRESENTATIVE RESULTS OF WATEGM CHEMICAL EQUILIBRIUM PROGRAM UTILIZING VARIOUS WATERS AS INPUT.

	pH	Ca	Mg	Na	K	Alka[2]	SO_4	TDS	SI[3]	SI	SI	SI	SI	%[4]	%	%
		← mg/L →							CALC	GYP	THEN	MIRA	EPSO	$CaSO_4^o$	$MgSO_4^o$	$NaSO_4^- + Na_2SO_4^o$
Initial effluent from FGD waste leached by spoil groundwater (P.V. 1-7)	8.5	384	515	800	130	172	5134	7927	+1.22	-0.6	-4.94	-4.14	-2.13	10	25	<1
Final effluent from FGD waste leached by spoil groundwater (P.V. 44-50)	7.6	413	163	250	28	328	2073	3800	+0.84	-0.12	-6.15	-5.35	-2.73	20	5	<1
Average of six samples of interstitial pore water squeezed from FGD waste	8.41	366	1172	2904	62	189	11672	21569	+0.72	-0.02	-3.75	-3.40	-1.74	4	24	7
Average of 2 piezometers near fly ash disposal sites[1]	7.38	378	218	3112	178	467	6811	9690	+0.25	-0.07	-3.90	-3.53	-2.47			
Average of 15 piezometers in spoils	6.75	284	153	325	19	670	1522	3400	-0.05	-0.30	-5.98	-5.71	-2.78			

[1] C-136 and C-139 (see Table 12).

[2] Alkalinity expressed as mg/L of $CaCO_3$.

[3] $SI = \log(IAP \cdot K_{eq.}^{-1})$ from WATEGM.

[4] % of total SO_4 (data for columns represent an average of three effluents).

estimating the contaminant yield capability because most of the total mass of each element is present in a relatively insoluble form.

Standard methods for determining the mineralogical composition of solids, such as x-ray diffraction and optical microscopic examination, are also only of limited use because these methods generally provide inadequate information on solid-phase components that are present as small percentages of the total volume of sample.

Although Figure 5 shows that effluent concentrations of calcium and sulfate decline with an increase in the cumulative number of pore volumes passed through the columns, this does not imply that disequilibrium exists with respect to gypsum. Computed saturation indices for gypsum indicate that the effluent is at or very near equilibrium with respect to gypsum throughout the entire period of effluent discharge (Table 9). The changing concentrations of calcium and sulfate at equilibrium are caused by the changes in concentrations of sodium and magnesium that are controlled by the rate at which these elements are released from the FGD waste to the flowing water. As less and less sodium and magnesium are released to the pore water, less sulfate exists in the form of sodium and magnesium ion pairs or complexes and therefore solubility of gypsum decreases.

Although important in the flowing column effluent, ion pairing and complexing is probably even more significant in the nonflowing squeezed porewater. Under nonflowing conditions the porewater can be expected to take on a higher concentration of sodium and magnesium sulfate salts than under the flowing conditions present in the columns. This will allow for greater gypsum dissolution and sulfate generation than under flowing conditions, as indicated by the very high equilibrium concentrations of sulfate in porewater squeezed from FGD waste (Table 9).

Mass balance calculations indicate that after passage of 50 pore volumes through the columns, less than 25% of the total sulfur had been removed from the columns. Based on the assumption that all of the sulfur represented in the elemental analysis of the FGD waste exists as calcium sulfate (anhydrite), it can be concluded that hundreds of pore volumes would have to be passed through the FGD waste in order to remove all of the available sulfate. After the soluble sulfate is removed, the waste will contribute only very low levels of dissolved solids to water in contact with the waste. A small fraction of the total sulfate in FGD waste or fly ash probably occurs as sodium or magnesium sulfate salts. Because these salts are very soluble, they would be quickly removed by flowing water. There appears to be no significant source of sulfate in FGD waste or fly ash other than sulfate salts.

EFFECTS ON GROUNDWATER AT THE CENTER SITE

Piezometers and Sampling Methods

Approximately one hundred and fifty piezometers have been installed within the waste-disposal area and unmined areas adjacent to the Center Mine (Figure 2). The piezometers were screened at various stratigraphic positions in both areas. The purpose of this instrumentation was two-fold: first, to gather baseline data on the occurrence, flow, and chemistry of groundwater in unmined and spoils settings; second, to evaluate the effects of FGD waste and fly ash on groundwater quality in disposal areas.

All piezometer test holes were drilled utilizing a rotary drilling rig. With few exceptions, the piezometer test holes were drilled using only air for circulation, thereby eliminating potential chemical contamination associated with the injection of fluids during drilling operations.

All piezometers consist of 5-cm-diameter PVC pipe with a 1.5 metre preslotted PVC screen. Washed sand was packed around the screened interval. Grout was then emplaced from the top of the sand pack to the surface. All piezometers were bailed at least twice prior to sampling. Sampling of the piezometers followed EPA recommendations. A more detailed discussion of well installation and sampling procedures can be found in Groenewold et al. (1979[2]). Water levels in all the piezometers were monitored on a monthly basis. These data were used to develop an interpretation of groundwater flow in the study area. In addition, single-well-response tests of selected piezometers were conducted to determine the hydraulic conductivities of the various units.

Groundwater Occurrence and Flow

As previously discussed, groundwater conditions in the Center area are typical of most active and proposed surface mining areas in the Northern Great Plains. A detailed discussion of hydrogeologic conditions in the Center area can be found in Groenewold et al. (1979[2]). The hydrologic regime is primarily a function of climate, stratigraphy, and topography. Most precipitation is lost through evapotranspiration. Recharge to the groundwater system occurs during spring runoff and occasionally during high intensity precipitation events.

Once infiltrating water moves below the rooting zone, the rates and directions of flow are largely determined by the lithology of the sediments (Figure 3). Movement of water through sand and lignite aquifers is generally lateral whereas movement through fine-textured Tertiary sediment and glacial till is generally vertical and downward (Moran et al., 1978[1], 1978a[7]).

The hydraulic conductivity of Tertiary sand commonly ranges from 10^{-5} to 10^{-6} m/sec; lignite generally ranges from 10^{-5} to 10^{-7} m/sec; fine-textured Tertiary sediments and glacial till range from 10^{-8} to 10^{-10} m/sec (Groenewold et al., 1979[2], Rehm et al., 1980[8]).

Groundwater of suitable quality for domestic use is not readily available in most areas of the Northern Great Plains. In western North Dakota, 60 to 70 percent of the small-yield domestic wells withdraw groundwater from lignite aquifers. The permeability of the lignite is largely due to fracturing, apparently a function of unloading. Thus, although as many as 10 or more lignites are present in many areas, typically only the near surface lignites are sufficiently fractured to constitute aquifers.

The near-surface lignites are presently proposed for extensive mining. Therefore, in many areas unmined lignite will be essentially in direct hydraulic contact with the basal portion of adjacent spoils areas. Thus the quality of water in spoils can potentially have a significant impact on groundwater in adjacent unmined areas. In addition, slow downward migration of waters from areas of spoils can be expected to affect the quality of groundwater in underlying units.

The basal portion of the spoils is often characterized by concentrations of blocky materials, particularly in the areas between spoils ridges (Winczewski, 1977[9]). This observation is substantiated by Van Voast et al., 1978[10]) who have determined that the basal portions of spoils at Colstrip and Decker Mines in Montana are characterized by greater average hydraulic conductivities than in other positions within the spoils. Their research also indicates that the mine spoils at Decker and Colstrip sites presently are at least as conductive of groundwater as the coal beds.

The movement of subsurface waters and permeability characteristics in the base of spoils areas in western North America have been addressed by various researchers (Alberta Environment, 1980[11], Davis and Rechard, 1977[12], Groenewold and Bailey, 1979[13], Groenewold and Rehm, 1980[14], Rahn, 1976[15], and Van Voast et al., 1978[10]). Hydraulic conductivity data from the basal portion of spoils in these areas show a wide range of values. Slug test values from spoils in Montana, Wyoming, and North Dakota range from 10^{-5} to 10^{-9} m/sec. Such a range of values seems reasonable given the high degree of structural, morphological, and textural variability within a given body of spoils.

Single-well-response tests were run on five piezometers screened in the base of the spoils at the Center mine. These data are presented in Table 10. The data indicate considerable variability in the hydraulic characteristics in the base of the spoils. Hydraulic conductivity values range from 10^{-6} m/sec, which is similar to lignite aquifers, to 10^{-9} m/sec, which is within the range of the fine-textured sediments.

Table 10. HYDRAULIC CONDUCTIVITY DATA FOR BASE OF SPOILS--CENTER MINE.

Piezometer No.	Hydraulic Conductivity (m/s)
C 9	1.4×10^{-9}
C10	5.9×10^{-6}
C16	2.8×10^{-9}
C17	5.0×10^{-7}
C23	3.1×10^{-6}

Groundwater instrumentation in the mined area indicates that the basal portion of spoils is saturated over most of the study site. Potentiometric head values in most of the piezometers in the spoils have shown a fairly consistent rise over the period of the study, suggesting that equilibrium conditions have yet to be established in the spoils. Potentiometric head data suggest that groundwater recharge is occurring through the spoils. This is verified by tritium and oxygen-18 data obtained from 22 piezometers in the spoils area. The oxygen-18 data indicate that the water in the base of the spoils has not been subjected to significant evaporation as would be expected if recharge were due to the lateral movement of water from the adjacent mine pit. The tritium data indicate that the water in the base of the spoils is enriched in tritium relative to groundwater in laterally adjacent and underlying undisturbed units, indicating vertical movement and recharge of this water through the spoils.

Groundwater Composition in Unmined Areas and in Spoils Without Waste

Chemical analyses of groundwater from undisturbed units and spoils at the Center Mine are shown in Table 11. Although considerable variability in water chemistry is apparent between the various undisturbed units, all contain water that is generically very similar. The major ions in solution in all the undisturbed units are sodium, calcium, bicarbonate, and sulfate. The Kinneman Creek and Hagel lignite beds contain Na, Ca-HCO_3, SO_4 to SO_4, HCO_3 type water. Total-dissolved solids concentrations in the Kinneman Creek and Hagel beds range from 843 to 1631 mg/L and 649 to 3874 mg/L, respectively. The nonlignitic sediments underlying the Hagel lignite bed or spoils are characterized by Na to Na, Ca-HCO_3, SO_4 to SO_4, HCO_3 type water. Total-dissolved-solids concentrations in these units range from 416 to 5051 mg/L.

Sulfate concentrations in the undisturbed units commonly exceed the recommended limit for drinking water of 250 mg/L by a factor of 2 to 4. The concentrations of iron and manganese in the undisturbed units vary considerably. Iron ranges from 20 to 1300 μg/L. The majority of the samples exqeed the recommended limit of 300 μg/L. Manganese concentrations range from 60 to 2750 μg/L, all in excess of the recommended limit of 50 μg/L. Cadmium levels are generally well below the maximum permissible concentration of 10 μg/L. Lead is typically well below the maximum permissible concentration of 50 μg/L. Selenium and arsenic are below the maximum permissible levels of 10 μg/L and 50 μg/L, respectively, in all samples.

The data presented in Table 11 indicate that the water in the base of the spoils is generally more highly mineralized, with respect to major ions, than groundwater in undisturbed units. Total-dissolved-solids concentrations in spoils at the Center Mine range from 1599 to 5338 mg/L. Sulfate concentrations range from 565 to 2520 mg/L. Concentrations of iron and manganese in spoils show a moderate increase over concentrations of these ions in undisturbed units. However, cadmium, lead, arsenic, and selenium in groundwater in spoils are within the same ranges of concentrations as in undisturbed units. All are well below the maximum permissible concentrations for drinking water.

Groundwater Composition in Waste Disposal Areas

Analyses of groundwater from piezometers in FGD waste and fly ash disposal areas at the Center Mine are shown in Table 12. Location of the screened interval relative to buried waste is noted for each piezometer.

FGD Waste. The FGD waste had been in place for approximately 12 months prior to sampling of the piezometers. At that time three piezometers, C94,

Table 11. CHEMICAL ANALYSES OF GROUNDWATER FROM SPOILS AND UNDISTURBED UNITS, CENTER MINE, CENTER, NORTH DAKOTA

	Field Temp °C	Field pH	Field Cond µmhos/cm	DO	TDS	Total Hardness	Ca	Mg	Na	K	HCO_3	CO_3	SO_4	Cl	NO_3	F
				←							mg/L					→
	SPOILS															
$\bar{x}$	9.7	6.7	3880.0	0.8	3375.6	---	348.0	232.4	361.0	15.6	1165.4	---	1454.9	8.1	0.6	0.1
s	4.2	0.3	1243.2	0.4	910.7	---	97.5	90.1	255.1	8.9	413.1	---	536.7	4.6	0.4	0.1
n	21	21	21	15	16	---	16	16	16	16	18	---	18	18	13	15
high	16.5	7.7	8000	3.5	5338.0	---	519.0	431.2	1118.0	34.0	2113.0	---	2520.0	21.0	1.8	0.3
low	2.0	6.3	1830	0.2	1599.0	---	190.0	103.5	89.5	7.5	363.6	---	565.8	2.3	0.1	0.0
	KINNEMAN CREEK BED															
$\bar{x}$	10.4	6.7	1655.6	0.4	1229.3	404.3	84.3	51.9	249.7	18.8	651.1	1.1	372.2	8.1	1.8	0.1
s	2.8	4.6	437.1	0.2	333.3	227.0	47.0	20.3	122.7	11.3	208.7	0.4	137.5	8.0	2.6	0.0
n	9	9	9	5	9	4	9	9	9	9	9	3	9	9	8	6
high	16.0	7.5	2200	0.7	1631.0	731.0	165.0	94.4	441.0	48.0	1021.1	1.4	642.7	21.0	6.5	0.1
low	6.5	6.0	975	0.2	838.0	221.0	29.0	29.0	115.0	11.4	334.3	2.6	232.6	0.5	0.1	0.05
	HAGEL BED															
$\bar{x}$	9.5	7.1	2264.3	0.4	1774.0	94.6	105.3	74.3	297.4	10.7	681.4	8.0	668.3	7.0	1.3	0.1
s	2.9	0.5	1041.8	0.5	1056.3	---	116.7	72.3	202.3	5.5	218.5	9.0	604.8	6.3	2.4	0.2
n	20	20	20	13	19	1	17	17	17	17	18	2	19	19	14	14
high	14.5	8.0	3750	1.9	3874.0	---	337.0	202.0	810.0	23.2	1098.0	14.3	2087.0	28.0	8.5	0.7
low	3.0	6.4	900	0.1	649.0	---	8.0	2.1	72.4	4.2	412.2	1.6	41.2	0.5	0.0	0.0
	SAND 5m BELOW HAGEL BED															
$\bar{x}$	9.2	7.4	3453.8	0.6	2644.8	---	153.2	80.4	567.7	10.2	835.0	---	1217.0	6.3	0.6	0.4
s	2.0	0.9	932.4	0.5	985.0	---	136.1	74.8	247.6	2.2	327.7	---	663.3	3.3	1.0	0.5
n	13	13	13	13	10	---	13	13	13	13	13	---	13	13	13	13
high	12.0	9.3	4400	1.6	4428.0	---	431.8	260.0	880.2	14.0	1288.2	---	2539.0	11.0	2.9	1.4
low	5.5	6.4	1900	0.0	1413.0	---	30.9	12.7	5.9	.74	106.8	---	415.0	2.0	0.0	0.0
	SILT 15m BELOW HAGEL BED															
$\bar{x}$	8.7	7.6	2946.5	0.5	1983.2	78.6	53.6	35.7	636.5	10.5	1154.7	12.8	694.7	11.1	2.2	0.7
s	2.6	0.5	1345.0	0.5	911.0	40.2	52.0	41.9	252.1	6.8	413.6	0.3	658.0	15.6	8.7	0.7
n	47	47	47	35	44	4	47	47	47	47	47	2	47	47	41	38
high	14.0	8.4	8000	2.7	5056.0	128.0	191.5	186.0	1311.0	36.4	1942.2	13.0	1642.6	107.2	55.8	2.3
low	3.0	5.9	600	0.1	416.0	43.3	3.5	2.8	106.5	3.5	248.0	12.6	140.9	0.5	0.0	0.0

Table 11 (continued). CHEMICAL ANALYSES OF GROUNDWATER FROM SPOILS AND UNDISTURBED UNITS, CENTER MINE, CENTER, NORTH DAKOTA

	SO_3	B	Alk.	Fe	Mn	Cd	Hg	Se	As	Pb	Ba	Cr	Co
	mg/L			µg/L									
	SPOILS												
$\bar{x}$	3.8	1.0	954.7	555.6	1951.2	1.9	0.3	1.6	3.7	4.9	92.8	7.3	13.3
s	1.8	1.2	317.0	557.7	1100.4	2.4	0.3	1.2	2.9	7.2	93.5	4.9	11.9
n	14	11	15	16	16	16	15	12	12	10	10	10	16
high	8.0	4.5	1732.0	1470.0	4070.0	8.8	1.3	4.8	10.1	22.1	348.5	18.1	46.1
low	1.5	0.5	298.0	20.0	2.8	0.5	0.0	0.0	0.4	0.0	10.4	1.9	1.8
	KINNEMAN CREEK BED												
$\bar{x}$	5.0	---	507.0	484.0	325.8	12.0	0.1	2.5	1.0	7.1	51.6	2.5	6.6
s	4.2	---	237.8	369.4	200.3	16.7	0.1	1.3	---	5.4	12.0	1.6	1.1
n	2	---	4	9	6	6	2	2	1	5	2	2	2
high	8.0	---	837.0	1300.0	610.0	34.0	0.2	3.4	---	13.8	60.0	3.6	7.4
low	2.0	---	274.0	45.1	95.0	0.2	0.0	1.6	---	2.0	43.1	1.4	5.8
	HAGEL BED												
$\bar{x}$	4.9	0.9	655.2	220.2	1072.7	1.2	0.4	0.8	2.8	16.9	159.6	6.0	8.9
s	6.1	0.9	399.2	156.4	1120.9	1.1	0.2	1.1	2.1	25.5	257.3	6.7	9.5
n	9	6	16	16	15	15	14	7	6	8	8	8	14
high	20.5	2.8	2040.0	580.0	3440.0	4.4	0.7	2.9	6.5	76.5	792.0	20.0	33.5
low	0.5	0.5	338.5	50.0	50.0	0.5	0.2	0.0	1.0	0.4	42.7	0.3	0.0
	SAND 5m BELOW HAGEL BED												
$\bar{x}$	5.9	3.5	684.5	236.2	660.0	5.0	0.3	1.6	8.1	11.5	54.2	13.0	21.6
s	8.6	2.3	268.7	222.9	765.6	7.1	0.3	1.8	13.4	12.2	22.7	11.4	18.7
n	13	9	13	13	13	13	13	9	9	7	7	7	13
high	31.0	7.2	1055.9	630.0	2930.0	23.5	0.8	6.1	37.6	38.8	79.5	36.8	66.1
low	1.0	0.5	87.5	20.0	120.0	0.5	0.0	0.0	0.9	3.2	7.2	2.9	0.0
	SILT 15m BELOW HAGEL BED												
$\bar{x}$	8.4	5.1	956.9	227.3	265.1	2.6	0.6	2.2	3.6	14.1	154.4	12.9	20.8
s	11.6	10.3	337.4	150.4	144.2	5.2	1.0	1.8	3.4	10.8	197.9	10.5	15.6
n	36	18	39	47	45	45	41	28	27	29	25	25	41
high	55.0	39.3	1592.0	640.0	770.0	25.2	4.6	6.1	14.0	39.0	758.0	52.6	83.9
low	1.0	0.5	203.0	30.0	49.0	0.2	0.0	0.0	0.0	0.9	31.6	2.1	0.2

C99, and C102, were beginning to show the influence of FGD waste leachate.

The groundwater in those piezometers is characterized by concentrations of sodium and sulfate that are greater than the average for those ions in unaffected spoil water by a factor of 2 to 3 (Table 12). Sodium concentrations range from 493 to 1845 mg/L; sulfate concentrations range from 3575 to 4917 mg/L in FGD waste-affected groundwater. Total-dissolved-solids concentrations in the FGD waste-affected groundwater range from 6564 to 7872 mg/L. The average total-dissolved-solids concentrations in spoil water at the Center Mine is 3375 mg/L.

Iron concentrations in FGD waste-affected groundwater are generally less than in unaffected spoil water. Manganese concentrations show approximately the same range as unaffected spoil water. The concentrations of cadmium, lead, arsenic, and selenium in FGD waste-affected groundwater are within the same ranges as spoil water in unaffected areas. None of these ions exceed permissible limits for drinking water. The pH of groundwater in FGD waste-affected areas ranges from 6.15 to 7.30.

These data indicate, as suggested by laboratory studies, that the major potential impact on groundwater of FGD waste from the Milton R. Young plant is the generation of exceptionally high concentrations of sulfate and other major ions. The generation of these ions in solution, as previously discussed, is largely dependent upon the dissolution of soluble salts. Therefore, these impacts can be minimized by the design and selection of disposal sites which isolate the FGD waste from infiltrating water.

Fly Ash. Piezometers showing effects of fly ash leachate at the Center Mine include four (C133, C136, C139, and C140) in areas where the fly ash has been in place for approximately 12 months. In addition, one piezometer (C10) is adjacent to an area where fly ash was buried in the spoil approximarely 6 to 7 years ago (Table 12).

The groundwater in these piezometers is characterized by exceptionally high concentrations of sodium and sulfate. Sodium concentrations range from 953 to 16,947 mg/L; sulfate concentrations range from 3213 to 36,820 mg/L. Total-dissolved-solids concentrations in fly ash-affected groundwater range from 5425 to 52,650 mg/L.

Iron and manganese concentrations vary considerably, ranging from 4 to 2590 mg/L and 10 to 6500 mg/L respectively. Cadmium concentrations in fly ash-affected groundwater are within the same range as unaffected spoils. Lead concentrations generally show a significant increase over concentrations in unaffected spoil, ranging from 8 to 236 µg/L.

Arsenic and selenium concentrations show extreme variability. Arsenic ranges from <1 to 613 µg/L in fly ash-affected groundwater. Selenium ranges from <1 to 800 µg/L. Noteworthy is the fact that both arsenic and selenium show highest concentrations in groundwater having high pH values. Molybdenum is extremely high in some of these samples, ranging from 218 to 38,460 µg/L. Molybdenum concentrations in lignite and sand aquifers in western North Dakota generally range from 10 to 30 µg/L. The pH in fly ash-affected groundwater at the Center Mine ranges from 6.95 to 12.1.

Table 12. GROUNDWATER ANALYSES FROM PIEZOMETERS IN AREA OF FGD WASTE AND FLY ASH DISPOSAL, CENTER MINE

Piezometer No.	Date	Field Temp.	Field pH	Field Cond.	TDS	Alk.	Ca	Mg	Na	K	HCO$_3$	SO$_4$	Cl	NO$_3$	F
					mg/L										
Piezometers Showing Influence of Fly Ash FGD Waste															
C 94	03-27-80	8.0	6.30	>8,000	7,872	163.4	195.3	615	900	55.0	199.4	4,917	10.7	-	<0.1
C 94	06-09-80	11.0	7.30	13,600	7,772	191	307	262	493	32	233.0	3,572	-	-	-
C 102	06-09-80	14.0	7.20	12,400	7,198	633	186	105	1,845	24	772.3	4,343	-	-	-
C 99	03-29-80	9.5	6.15	8,000	7,220	553.4	215.5	530	960	24.2	675.2	4,111	22.0	-	0.73
C 99	06-09-80	14.0	7.25	10,000	6,564	645	393	408	1,050	24	786.9	3,811	-	-	-
Piezometers Showing Influence of Fly Ash															
C 10	11-10-78	6.0	12.10	>8,000	5,425	485	32.8	0.2	1,550	198.4	591.7	3,213	27.0	0.60	1.4
C 10	07-06-78	11.5	11.74	>8,000	5,548	715.2	28.2	0.1	1,340	9.7	872.5	3,015.5	19.8	1.13	1.7
C 10	02-24-79	4.5	11.85	>8,000	5,715	500	30.8	1.3	1,517	198.3	610.0	3,607.3	26.5	0.13	1.65
C 133	06-19-80	15.0	8.75	21,600	12,270	182.7	780.7	39.9	3,364	500	209.3	8,893	11.9	0.15	0.16
C 136	05-20-80	13.0	7.80	29,000	12,420	364.1	240	123.0	5,119	280	432.3	9,265	13.0	0.10	0.49
C 136	07-28-80	13.2	7.9	6,500	10,690	381.6	220.4	131.9	2,963	287.6	460.3	6,723.2	11.3	0.21	0.66
C 139	05-20-80	13.0	6.95	13,600	6,960	569	515	312.9	1,303	75	688.1	4,357	9.0	0.40	0.23
C 139	07-28-80	12	7.25	7,800	6,720	578.8	524.7	292.8	953	35.4	698.6	3,428	9.8	0.24	0.32
C 140	06-19-80	14.5	11.55	99,200	52,650	534.9	130.4	20.6	13,573	2930	24.5	36,820	74.9	0.34	0.84
C 140	07-28-80	17.0	11.62	>8,000	50,110	684.6	97.7	19.6	16,947	-	16.7	31,045	57.7	0.23	2.34

Position of Screened Interval

C 94	In spoils, offset from FGD waste approximately 10 m
C 99	In base of spoils, 11 m below FGD waste
C 102	In base of spoils, slight offset from FGD waste
C 10	In base of spoils, offset from fly ash unknown distance
C 133	In base of spoils, offset 20-25 m and 10 m below fly ash
C 136	In base of spoils, 11 m below fly ash
C 139	In base of spoils, 7.5 m below fly ash
C 140	In spoils, immediately below fly ash

Table 12 (continued). GROUNDWATER ANALYSES FROM PIEZOMETERS IN AREAS OF FGD WASTE AND FLY ASH DISPOSAL, CENTER MINE

Piezometer No.	Fe	Mn	Cd	Hg	Se	As	Pb	Mo
	µg/L							
Piezometers Showing Influence of Fly Ash FGD Waste								
C 94	130	4,410	1.8	<0.3	<2	3.0	-	-
C 94	20	-	<1	<0.3	<2	<2	7.3	-
C 102	90	-	3.3	<0.3	5.6	<2	29	-
C 99	50	1,740	1.4	0.6	2	2	-	-
C 99	250	-	<1	<0.3	<2	<2	7.4	-
Piezometers Showing Influence of Fly Ash								
C 10	30	24	3.6	0.7	13.6	204.6	58.0	-
C 10	20	10	<1	0.8	3.8	171.8	33.1	-
C 10	110	20	4.7	0.3	8.2	97.8	-	-
C 133	140.7	-	3.75	-	-	13.3	7.67	17,019
C 136	4	-	0.82	-	-	22.6	236.0	4,500
C 136	308	495	0.24	-	0.56	35.3	28.1	5,466
C 139	2,080	-	0.39	-	-	0.3	60.0	640
C 139	2,590	6,500	0.9	-	0.56	6.2	16.1	218
C 140	21.8	-	0.78	-	800	550	34.1	23,700
C 140	1,376	271	4.3	-	760	613	-	38,460

These data indicate that fly ash from the Milton R. Young plant can potentially cause groundwater to acquire exceptionally high concentrations of sulfate and other major ions. In addition, the fly ash has the potential to severely degrade groundwater with respect to arsenic selenium, molybdenun, and possibly lead. Proper disposal of fly ash is therefore of critical concern. The selection and design of disposal sites for fly ash will require extreme caution and a detailed knowledge of the hydrogeologic conditions in proposed disposal areas.

SUMMARY OF CONCLUSIONS

Chemical analyses of absorber tower liquor, supernatant solution from transportation vessels and from batch shaker tests, pore water extracted by squeezing of FGD waste samples, and effluent from column experiments all indicate that waters in contact with FGD waste from the Milton R. Young power plant are characterized by sulfate concentrations sufficiently high to make the water unfit for drinking and for most agricultural uses. These waters have pH values that are not much above the pH of natural groundwater in the Center area. Water in contact with FGD waste that has not been previously leached acquires fluoride concentrations that are commonly greater by a factor of 2 to 4 than the limit specified for drinking water. Lead is generally greater than the drinking water limit by a factor of 1.5 to 3, and cadmium by a factor of 1.5 to 5. Insufficient data are available to draw general conclusions for selenium; preliminary results suggest that this element will exceed the drinking water limit by a factor less than 2. When water continuously passes through FGD waste, the concentrations of fluoride, lead, and cadmium generally decline to levels near or below the drinking water limits but sulfate remains above the limit by a factor of 5 to 10.

Laboratory studies indicate that water in contact with fly ash becomes very alkaline with pH generally between 11 and 13, and commonly acquires concentrations of arsenic and selenium that exceed the drinking water limits by factors as large as 15 and 35 respectively. In terms of major ion composition, the fly ash water should be similar to FGD waste water in that it should be characterized by high concentrations of sulfate, sodium, magnesium, and calcium.

The natural groundwater in coal-mining areas in western North Dakota commonly has sulfate concentrations that exceed the recommended limit for drinking water. Groundwater in spoil in areas of reclaimed land commonly has exceptionally high sulfate concentrations and therefore is sometimes unsuitable for domestic or agricultural use. FGD waste and fly ash can cause exceptional changes in the quality of groundwater because of the input of toxic constituents such as As, Se, F, Pb, and Cd to groundwater and in this regard fly ash is a considerably greater hazard to groundwater than FGD waste.

Although the fly ash and FGD waste contain about the same quantities of arsenic and selenium, only the fly ash imparts high concentrations of these elements to water. This is attributed to the effect of pH; laboratory experiments indicate that the high pH caused by the fly ash may be responsible for the high arsenic and selenium concentrations. At pH levels in the normal groundwater range, the concentrations of these elements are rarely high. Although it is known that high aresnic and selenium concentrations occur at high pH and rarely at lower pH, the specific chemical effect of pH is not understood at present. It appears that desorption of these elements from the surfaces of solids caused by the increase in hydroxyl ions may be a major factor. The fact that water in contact with fly ash attains a very high pH and water in contact with FGD waste does not is due to the relatively higher content of oxides of calcium, magnesium, and sodium in the fly ash. The oxides undergo spontaneous hydrolysis to produce a large deficiency of hydrogen ions and thus the high pH.

Calculations of saturation indices indicate that sulfate concentrations of water in contact with fly ash and FGD waste are limited by the solubility of gypsum and that gypsum solubility is the main factor limiting the levels to which total dissolved solids and electrical conductance can rise. Although pore waters extracted from and the effluent from FGD waste columns are at or close to equilibrium with respect to gypsum, the concentration of sulfate, calcium, magnesium, sodium, and total dissolved solids vary considerably. The equilibrium concentrations of calcium and sulfate can vary, because sodium and magnesium form pairs and complexes with sulfate that can comprise a large percentage of the total sulfate when sodium and magnesium concentrations are high. The amounts of sodium and magnesium in soluble form are therefore important controls on the levels to which sulfate, calcium, and total dissolved solids can rise. As water passes through FGD waste, the leachable sodium and magnesium are removed more rapidly than calcium and sulfate and therefore the influence of ion pairing and complexing decreases with a corresponding decrease in the equilibrium sulfate concentrations to values generally in the range of 1400 to 2000 mg/L.

The ability of fly ash and scrubber waste to cause groundwater to acquire exceptionally high concentrations of sulfate and major cations was confirmed by the results of sampling of piezometers installed in spoil at the Center site where these wastes have been buried. The occurrence of groundwater with pH values in the range of 11 to 13 and high concentrations of arsenic selenium, and molybdenum provided confirmation of the conclusions that fly ash has the potential to cause severe degradation of groundwater in reclaimed land. Arsenic and selenium in high pH groundwater are contaminants with exceptionally adverse hydrogeologic characteristics because they exist in solution as very soluble anions that may not undergo much geochemical attenuation as they are transported along groundwater flow paths. The mobility of these anionic contaminants in groundwater in spoil and adjacent zones is a subject of current research.

Although the primary purpose of fly-ash-type flue gas desulfurization systems is to reduce atmospheric contamination, the results of this study show that a secondary benefit of this method of flue-gas processing is the conversion of fly ash from a form that can cause groundwater to acquire severe toxicity because of high aresenic and selenium levels to a waste form that causes increased sulfate concentrations but generally no significant increases in the more toxic elements.

Burial of FGD waste in mined areas offers an effective means of disposing of this waste, providing that the waste is placed in favorable locations in the mined area and provided that the selective placement of the waste and spoil is accomplished with appropriate consideration for the hydrologic and geochemical nature of the system.

Placement of these wastes in V-notch settings appears to be highly preferable to pit bottom disposal in most areas. V-notch areas are commonly above the postmining water table, and thus offer much less opportunity for the dissolution and leaching of soluble salts present in these waste products. Ongoing research will better evaluate the suitability of various postmining landscape positions for the disposal of fly ash and fly ash FGD waste.

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Environmental Compatability and Engineering Feasibility for Utilization of FGD Waste in Artificial Fishing Reef Construction

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Abstract

A multidisciplinary team at the State University of New York at Stony Brook, in collaboration with IU Conversion Systems, Inc. (Horsham, Pa.), is assessing the feasibility of using blocks of waste materials from coal-fired power plants for underwater construction of artificial fishing reefs. Experiments conducted over the past three years in the laboratory and in the sea have suggested that coal waste blocks are environmentally acceptable in the ocean.

On September 12, 1980, a 500 ton reef was constructed in the Atlantic, south of Long Island, from 18,000 solid blocks 8" x 8" x 16" of stabilized fly ash and FGD sludge obtained from coal burning power plants located in Ohio and Indiana. The reef blocks were fabricated at a commercial concrete block plant in Pennsylvania using automatic block making equipment. The ratio (dry weight) of fly ash to FGD sludge was 1.5:1 and 3:1.

In preparation for fabrication of the 500 tons of reef blocks, a combination of different coal waste mixes, stabilization additives, and curing procedures were screened to develop candidate mix designs. Experiments were made at the research facilities of the Besser Company in Alpena, Michigan to develop methods by which coal wastes could successfully be formed into blocks by block making machines.

The demonstration disposal reef will be monitored for three to four years to assess environmental impacts which may occur and to measure the development of the biological communities which will be associated with the reef.

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A multidisciplinary team at the Marine Sciences Research Center, (MSRC) and the Materials Sciences Laboratory (MSL) of the State University of New York at Stony Brook, is investigating the long term interactions in the ocean of solid blocks made of combustion wastes from coal-fired power plants. The objective is to assess the feasibility of using blocks of the coal waste materials for underwater construction of artificial fishing reefs. Results of variety of experiments conducted over the past three years, first in the laboratory and later in the sea, have suggested that coal waste blocks may be environmentally acceptable in the ocean. The program has recently built a demonstration pilot reef from 18,000 waste blocks in the Atlantic off Long Island, N.Y. The reef will be studied in situ for three years.

There is urgency to convert from oil to coal burning at northeastern power plants and conversion has begun. An important obstacle to utilizing coal is the large volume of coal combustion products, the FGD sludge and fly ash, which require disposal.

The dumping of either the untreated scrubber sludge or fly ash in the sea would be quite unacceptable, probably having deleterious environmental effects. However, IU Conversion Systems, Inc., Pa., has developed a marketable stabilized coal waste by combining the scrubber filtercake with the fly ash. Basically this system treats sludge and fly ash with additives and cementitious reactions convert the mix to a stable material that can range from a clay-like substance to hard blocks. The stabilization reactions in the formation of the blocks are similar to the pozzolanic reactions which occur in the forming of concrete. This stabilized mixture is being used to fabricate blocks for artificial reef building. The bottom ash can be included as an aggregate.

Our research has been directed at determining the physical and chemical characteristics of the stabilized blocks of coal waste in sea water systems and what environmental effects, if any, the blocks might have. In particular, we are looking at how well the blocks serve as substrates for settlement and colonization by the plants and animals which associate with reefs.

Laboratory Investigations

Work began four years ago with laboratory studies funded by the New York State Energy Research & Development Authority, New York State Sea Grant Institute and the Link Foundation and performed by MSRC at Stony Brook on blocks provided by IU Conversions Systems, Inc., Pa.

Small test blocks were studied in the laboratory to characterize chemical and mineralogical composition, and to determine their physical and chemical properties. In their physical properties, coal waste blocks have considerable similarities to concrete but do not have the yield strength of concrete and are more porous and permeable. The bulk density of the blocks is about 80% that of concrete, due to the lighter fly ash used and the absence of high density aggregate materials.

Compressive strength values of coal waste blocks are only 25 to 50% that of concrete, but in seawater some of the blocks continued to cure, increasing in density and in strength up to 50% in four months. These changes appear to be associated with the slow formation of a new chemical phase within the blocks.

Several studies have considered leaching characteristics of coal waste blocks. Calcium and sulfate at first leach fairly rapidly from test blocks in tanks of seawater. But as leaching continues the rate of release of these major components decreases as the concentrations of the more soluble phases in the outermost layers of the blocks decrease. Leachates are also analyzed for trace elements such as iron, nickel, copper and mercury. Some elements show an initial increase in the seawater in the first days of exposure but after a few days were again taken up into the blocks, other elements did not dissolve at all. The behavior of dissolved trace elements was probably due to desorption-absorption processes, the trace elements remaining associated with the fine materials such as fly ash in the blocks.

Using procedures recommended by the U.S. Environmental Protection Agency, bioassays were performed on block elutriates in seawater at relatively high concentrations to provide information on material toxicity. Using sand shrimp, developing fish eggs, and newly hatched fish larvae (sensitive early life stages), elutriates appeared to have no effect upon viability. Other assays were made with a unicellular plant, a marine diatom. Measurements of the daily growth, or rate of reproduction, and of photosynthesis by the plant cells indicated that the elutriates did not inhibit growth.

Inshore Habitats

The first investigations of coal waste blocks in the sea were made in estuarine bay off Long Island Sound in about 18 feet of water. Several 1 ft^3 waste blocks were stacked into separate small habitats, "mini-reef" formations; one reef of blocks high in calcium sulfite. Concrete blocks were used for a control formation and a number of large natural rocks were also stacked nearby. The "mini-reefs" have been periodically examined for biological colonization and photographed by SCUBA divers in a series of field experiments over a span of three years. At intervals, test-blocks and encrusting organisms have been removed for laboratory analyses.

In the sea, the blocks have retained their physical integrity and, although there were strong tidal flows, block edges remain sharp with little erosion. Test blocks removed from the site showed that the strength of the blocks was maintained over extended periods. The blocks high in calcium sulfite increased progressively in compressive strength from 320 to 730 psi during one year on the sea-bed.

We have found no adverse environmental effects resulting from the placement of the waste blocks. Seaweeds and animals have attached themselves and overgrown the waste blocks, as they have also on the concrete blocks and the rocks placed at the site. There appears to have developed a diverse, productive community of reef organisms on all of the blocks. At first, there were some differences in the type of settlement on the different materials, but as the blocks became more heavily overgrown and finally encapsulated by plants and encrusting animals, the initial differences in colonization between the coal waste blocks and concrete began to disappear. After a year, differences were no longer evident.

Because the coal waste materials contain trace amounts of potentially toxic elements, samples of organisms growing on the blocks were removed by SCUBA divers for trace element microchemical analysis. Samples were analyzed for Cu, Cr, Zn, Pb, Cd, Hg, Ag, Se and As using atomic absorption spectrophotometry and other methods. The collections and analyses were repeated on five occasions over two years. In no instance was there evidence of elevated levels for any of the trace metals in the biomass collected.

The continuing laboratory and field studies strongly suggest that blocks of stabilized coal combustion wastes may be environmentally acceptable in the sea. An initial economic survey indicated that the concept of block disposal in the ocean offered savings relative to land disposal of wastes from a power plant situated on the coast or an estuary. The research has advanced to a new stage, moving to establish a pilot project artificial fishing reef in the open Atlantic.

Demonstration Reef in Atlantic

The program has now established the larger demonstration project, building a pilot reef with 500 tons of coal waste blocks, which were made by IU Conversion Systems, Inc. using methods developed by the program. The blocks have been placed 3 miles south of Long Island at a depth of about 70 ft in the New York Bight. The program is being funded by U.S. Environmental Protection Agency and U.S. Department of Energy, by the Electric Power Research Institute, and by New York State Energy Research and Development Authority and the Power Authority of the State of New York.

In preparation for fabrication of the 500 tons of reef blocks, different coal waste mixes, stabilization additives, and curing procedures were screened to develop candidate mix designs. Large scale experiments in block manufacture were carried out in Ohio where 1 yd^3 blocks, weighing about 1 ton, were built. Subsequent assessment of these experiments suggested that it might be cheaper and faster to produce smaller blocks (weighing about 60 lbs.) using conventional concrete block-making machinery. This was confirmed in another large scale investigation which we made at the research facilities of the Besser Co. in Alpena, Michigan, where methods were developed to form coal wastes into blocks with block machines. This technology was successfully transferred to the commercial factory this summer by demonstration experiments at the Fizzano Bros. concrete block factory equipment was used--demonstrating engineering feasibility. In the block making process FGD sludge, fly ash and additives are thoroughly mixed and run into the hopper of a block machine; strong vibration is used both to feed the material into steel molds and to compact the molded blocks on pallets. The pallets of green blocks are loaded on racks holding 540 blocks and cured for a day in steam kilns. Cured blocks are unracked, depalletized and stacked for handling as cubes of up to 144 interlocked blocks by a cubing machine, Figure 1. Block making is fast, a block machine can form more than 1,500 concrete blocks per hour and our calculations suggest that single machine working 3 shifts per day could process the wastes from a 500 MW plant. By employing steam kilns, curing is accelerated and greater block strength can be achieved in 24 hours than in 28 days of curing at air temperature. Accelerated curing allows immediate handling by automated machines and cured blocks may be rapidly disposed, minimizing storage space.

For the full scale manufacture of 500 tons of reef blocks, coal wastes were trucked from the Columbus & Southern Ohio Electric Co., 800 MW power plant at Conesville, Ohio and from the Indiana Power and Light Co. 530 MW plant at Petersburg, Indiana. Both are modern plants, Conesville employing lime scrubbers, Petersburg has limestone. The blocks were made at the factories of Fizzano Bros., and at York Construction Corp. in Harrisburg, Pa. The mixes used had fly ash to scrubber sludge ratios of 3:1 for Conesville waste and 1.5:1 for Petersburg waste. About 18,000 blocks were made, they were loaded on an oceangoing, bottom opening dump barge, and released at the Atlantic demonstration project site on 12 September 1980.

Prior to placing the reef, we made a series of baseline oceanographic cruises to characterize the project site and surrounding areas. The artificial reef will now be monitored for three or four years to assess environmental impacts which may occur and to measure the development of the biological communities which will be associated with the reef. Throughout the study extensive testing will be performed on blocks periodically removed from the demonstration reef to evaluate their acceptability as materials for fishing reef construction from physical, chemical, and biological perspectives. Other tests will be made by SCUBA divers on blocks remaining in the sea, including ultrasonic sensing of internal structural change. We hope that if this extended program of testing and oceanographic monitoring finds the blocks to be environmentally acceptable in the ocean, and without adverse effects, we may have demonstrated an alternative for the disposal of coal wastes which can also carry benefits for man and the marine environment.

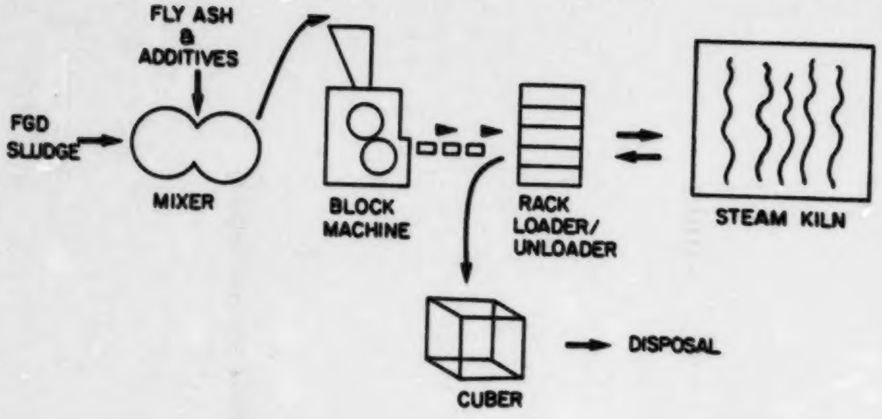
FLY ASH
&
ADDITIVES
FGD
SLUDGE
MIXER
BLOCK
MACHINE
RACK
LOADER/
UNLOADER
STEAM KILN
CUBER
DISPOSAL

Fig.1 Simplified schematic of block processing

GOVERNMENT PROCUREMENT OF CEMENT AND CONCRETE CONTAINING FLY ASH

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U.S. Environmental Protection Agency

The Resource Conservation and Recovery Act (RCRA) has two major objectives: 1) the protection of human health and the environment, and 2) the conservation of valuable material and energy resources. I think it is clear to everyone in this audience that we at EPA have been concentrating our efforts primarily on accomplishing the first of these goals, that is, protection of the public health. We believe this is only proper in view of the serious hazardous waste disposal problems which have come to light in the past few years. During the short term, at least, we must be concerned with providing adequate protection for the land, rivers and lakes, and groundwater supplies upon which we all depend to sustain our lives. Earlier this year we issued the first portions of hazardous waste regulations under RCRA. In those regulations, EPA granted a temporary deferral to fly ash, bottom ash, boiler slag, and scrubber sludge resulting from the combustion of fossil fuels. By this deferral, these waste streams are not considered to be hazardous, and will not be regulated as hazardous unless EPA finds evidence to indicate that they should be, after the completion of extensive studies during the next one and a half years. Initial indications are that none of these four waste streams are hazardous, according to EPA characteristics for determining hazard.

We do not intend to lose sight of the second major goal of RCRA, however. In passing the law, Congress stressed its support for its resource recovery objective by citing the following facts: millions of tons of recoverable material are needlessly disposed of annually; methods exist to separate and reuse these materials; and recovery and conservation of such materials can redue the national balance of payments deficit. The desire to increase resource recovery set the stage for inclusion of Section 6002 in RCRA, which is the topic of my presentation. This section of RCRA provides one of the few tools EPA has for influencing recovery and reuse of waste materials. We have selected fly ash used in cement and concrete as the very first product area which we are addressing under Section 6002.

First, let me give you a little background on Section 6002, which is entitled "Federal Procurement." It has one basic mandate--procuring agencies must purchase items containing the maximum practicable amount of recovered materials. This must be done, however, in keeping with what we refer to as the four "reasonables." That is, recycled products must be reasonably equivalent in technical performance to the virgin product. It must be available at a reasonable price, as compared to the typically virgin product which is normally purchased. Reasonable levels of competition must be maintained. Finally, the product must be reasonably available, in terms of delivery time and geographic locations. We firmly believe that fly ash, when used either in blended cement or as an admixture in concrete, can meet all of these "reasonables" in the majority of government procurements.

Especially in the case of cement and concrete, it is crucial to recognize that the term procuring agency is not just limited to direct purchases made by Federal agencies. Congress specifically applied this provision not only to the Federal agencies, but also to State and local governments, grantees, and contractors which are using Federal funds for procurement. For example, we consider the Federal Highway Trust Fund monies to fall under the requirements of Section 6002. Thus, the effect can indeed be very extensive.

Under RCRA, EPA is given responsibility for issuing guidelines to assist agencies in complying with purchasing responsibilities mandated to them by Congress. Basically, Section 6002 is mandatory, but the guidelines are voluntary. Procuring agencies will have to take action on the products designated in the guidelines, but not necessarily in accordance with our recommendations. Thus, we plan to provide the type of information which procurement officers have indicated they need--availability, costs, relative performance, percentage of recovered material, certification procedures, etc. If procuring agencies do adopt our guidelines, we feel they will be complying with the intent of RCRA.

Soon after EPA undertook this effort, it became clear that this type of "procurement" information could not be developed for all 50,000 Federal specifications and standards. We decided to concentrate only on those items which demonstrated the highest potential for resource recovery through the use of the Federal procurement "tool." Criteria were developed to aid in that selection process. These criteria are: 1) the waste material must constitute a significant solid waste management problem, due either to volume, degree of hazard, or difficulties posed by disposal; 2) eonomic methods of separation and recovery must exist; 3) the material must have technically proven uses; and 4) Federal purchasing power for the final product must be of sufficient size to be able to affect raw materials utilization policies on the part of manufacturers.

With these criteria in mind, EPA has chosen initially to issue four major product guidelines under Section 6002: recycled paper products, construction products containing recovered materials, composted sewage sludge as soil conditioner and fertilizer, and cement and concrete containing fly ash.

You may begin to see from the criteria we have established and the product areas initially selected that we are approaching this issue from the solid waste perspective. In other words, EPA views recovery and reuse of materials primarily as a solid waste management alternative -- and as a more desirable alternative than disposal.

Under our scenario, fly ash becomes a very attractive waste material for increased usage through the Federal procurement mechanism. Not only does its use in cement and concrete reduce the amount of waste requiring disposal, but its reuse in these applications has tremendous potential for both energy conservation and cost savings. This now brings us to the main topic of my discussion, which is Federal procurement of either cement or concrete which contains fly ash.

As some of you already know, EPA circulated a preliminary draft of our fly ash guideline for public comment in January. As a result of public comments and meetings which have been held with the affected Federal agencies, we have made some revisions to that earlier draft. A revised guideline package is currently winding its way through the EPA regulation review process, which includes review by all offices within the Agency as well as by all of EPA's regional offices. We are expecting to formally propose the guideline in the Federal Register by the end of 11/80 .

The formal proposal opens up an official 60-day public comment period, with a public hearing to be held around the fiftieth day. It is during this time that individuals and organizations are able to make their views known for the public record. We will evaluate the comments and concerns expressed during this period and make any appropriate changes to the guideline. We would anticipate publishing a final guideline around 7/81. We strongly urge you, who are so vitally concerned with fly ash, to send us your comments during this period.

With all of this discussion about the guideline, I really should mention what it contains. The main recommendation which we make is that agencies allow either blended cement or concrete containing fly ash to be bid as an alternate material to portland cement or concrete, except where the application would be technically inappropriate. The bidding mechanism will then determine whether cement and concrete containing fly ash is in fact available and at competitive prices. Award would be made to the lowest priced responsible bidder, regardless of whether fly ash is used. However, with the cost advantages of using fly ash, allowing it to be bid as an alternate should greatly increase its use. We further recommend that when, and if, a procuring agency becomes satisfied that cement and concrete containing fly ash is indeed available in a reasonable period of time, at a price competitive with that of portland cement or concrete, only bids for cement or concrete containing fly ash should be solicited. EPA feels constrained to include this provision in the guideline because of the strict way in which Section 6002 is written.

We had originally considered mandating the use of fly ash in all cases except where technically inappropriate. However, there was a strong negative reaction to this recommendation from both the inter-agency work group and preproposal commenters. Even persons who should have favored this approach -- often the distributors and users of fly ash -- believed requiring the use of fly ash to be unnecessary and counterproductive. These organizations pointed out that the cost savings from the use of fly ash would allow suppliers of cement and concrete containing fly ash to compete favorably with portland cement and concrete suppliers in equal bidding situations.

The three major disadvantages of requiring the use of fly ash are:

(i) the burden of determining reasonable levels of price, availability, performance and competition falls completely upon the contracting officers, prior to issuance of solicitations. Procuring agencies have indicated they do not have the expertise or resources to fulfill this requirement.

(ii) many potential suppliers lack experience in the use of fly ash. Mandating its use in all cases at the present time could result in its use by those currently unfamiliar with the material. Also, mandating the use of fly ash could encourage the use of low quality fly ash until fly ash of sufficient quality is available, resulting in possible deleterious performance of the concrete.

(iii) requiring the sole use of fly ash to the exclusion of other materials could prevent current efforts to promote use of other recovered materials in cement and concrete, particularly blast furnace slag from iron production.

Another issue of significant concern to reviewers has been quality control. Quality control and assured performance are difficult problems with any concrete mix design. The use of fly ash complicates the problem since the quality of the ash must be tested and adjustments must be made in the mix design. However, ample specifications and testing methods exist for ensuring the quality of the fly ash and the resulting cement or concrete containing fly ash. We definitely feel that only those ashes which, as a minimum, meet ASTM specifications should be used in cement and concrete. In addition, let us not forget that when the use of fly ash is allowed, cement and concrete suppliers are clearly responsible both for the quality of the ingredients of their product and for meeting appropriate performance standards, just as they are for portland cement concrete.

Another issue of concern to us in the development of this guideline has been radioactivity. Work group members and commenters were concerned with any potential health hazards which might be posed by the use of fly ash in habitable structures. The concern arises from the fact that fly ash contains radionuclides, as do almost all materials. Recent investigations indicate that use of some low level radioactive materials (e.g., phosphate slag, uranium mill tailings) in and around houses may cause an increased cancer health risk to the occupants. Because some fly ash exhibits very low levels of radioactivity, commenters suggested that similar risks may be present with fly ash.

Although current data are limited, there is no evidence to indicate that fly ash used in habitable structures would cause any health risk above that associated with the product it would partially replace, i.e., portland cement. Levels of radium-226 concentration in fly ash do not appear to vary significantly from those of portland cement. In addition, initial indications are that the physical characteristics of fly ash particles may in fact reduce radon emanation from fly ash - the real concern - to an extremely low level. The Office of Solid Waste is actively working with the National Ash Association (NAA) in an attempt to obtain additional data to completely resolve this issue.

The amount of fly ash generated has risen significantly over the years, and is expected to increase even more rapidly with construction of additional coal burning power plants during the 1980's. The quantity is estimated to be 70-80 million tons per year by 1985. What we want to affect is how much of that fly ash will be recovered and utilized. Extensive efforts by industry and the NAA have served to achieve an admirable 17.5 percent reuse level in 1978. However, we will need to see an increase here if a dent is to be made in the additional tonnages projected. We hope that this program will be but one of the driving forces behind it.

Session 6: DRY SCRUBBING

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SPRAY DRYER FGD: TECHNICAL REVIEW AND ECONOMIC ASSESSMENT

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ABSTRACT

This paper summarizes the results of an EPA-funded study of dry scrubbing technology and economics. The relative economics of a generic lime spray dryer process and a limestone scrubbing process were compared for three coal applications: a low-sulfur western, and low- and high-sulfur eastern coals. The cost estimates are based on recently updated TVA design and economic premises and include all of the processing required to meet the 1979 new source performance standards (NSPS) for both particulate matter and SO_2 and to dispose of the resulting FGD waste in an environmentally acceptable manner.

The resulting preliminary economics for all three coal applications for both the generic lime spray dryer process and the limestone scrubbing process are included. Sensitivity of the resulting annual revenue requirements to the delivered raw material cost and the raw material stoichiometry are also included.

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SPRAY DRYER FGD: TECHNICAL REVIEW AND ECONOMIC ASSESSMENT

INTRODUCTION

Dry scrubbing FGD technology, particularly that phase in which spray dryers are used, is currently receiving a considerable amount of attention in the utility industry. An alkaline solution or slurry is atomized in the flue gas and evaporates to dryness while reacting with the SO_x and HCl. The resulting reaction products are collected, along with fly ash, and disposed of as a dry waste. The method has several potential advantages over wet scrubbing FGD because it eliminates the complexity and operating problems associated with the large volume of scrubbing liquid used in wet scrubbing as well as liquid waste disposal problems. Conversely, high removal efficiencies are more difficult to attain and a highly reactive (i.e., nonlimestone) absorbent must be used.

In the past few years a number of companies and consortia have entered the spray dryer FGD field with pilot studies and several have contracted to build commercial units. Most of the pilot studies and all of the utility units are for low-sulfur western coal applications, where removal efficiencies and absorbent consumption are usually lower than with high-sulfur coals and, in some cases, the high alkalinity of the fly ash can supplement the absorbent. The rapid growth of spray dryer FGD is in part a result of the increasing use of western coal but its rapid growth also owes its derivation to the proven industrial technologies of spray drying and particulate matter collection. The development of fabric filter fly ash collection for utility use has been particularly advantageous. In many cases, companies active in spray dryer FGD have backgrounds in these technologies.

Spray dryer FGD is related to, and in some aspects evolved from, earlier efforts in dry injection of absorbents. Although many of these studies were disappointing in terms of SO_x removal efficiency, absorbent utilization, and availability of economical absorbents, interest in such uncomplicated approaches to FGD has continued. The development of spray dryer FGD, its general technical aspects, and its status through the early months of 1980 are discussed as a portion of this paper. Most of the information was developed from interviews with vendors and plant visits. In addition, the history and current status of dry injection processes are reviewed.

An important question of spray dryer FGD is its economics in relation to limestone wet scrubbing processes, in which the absorbent is less expensive. Although various vendor economic comparisons have been made, there have been no previous independent economic comparisons applicable to general utility applications. As the key feature of this paper, an economic comparison of a generic lime spray dryer process and a limestone scrubbing process is made for low-sulfur western coal and high- and low-sulfur eastern coal applications. A generic soda ash spray dryer process is also evaluated for the low-sulfur western coal case.

BACKGROUND

Dry absorption of SO_2 received considerable attention during the early 1970's because it appeared to have several technical advantages over wet scrubbing (1). Of the potential absorbents only sodium-based materials were found to be sufficiently reactive and, of these, nahcolite proved the most effective. However, owing to economic and environmental considerations, commercial mining of the large nahcolite reserves in Colorado has not occurred and questions about the widespread availability of nahcolite in the future forced a search for other absorbents. Because other more readily available raw materials are either too expensive or not sufficiently reactive, development of dry absorption FGD slowed in the mid-1970's and primary emphasis focused on the technology to make the readily available reactants more reactive without losing the potentially significant advantages of dry FGD. This search (along with the simultaneous development of the regenerable aqueous carbonate process) led to the development of spray dryer FGD.

Recently interest in dry absorption has resurfaced with several pilot-plant programs (2,3). For absorbents, nahcolite is still the primary focus but trona is also being evaluated. Trona unlike nahcolite has the advantage that it is already being mined in commercial quantities for the production of soda ash. Although early results appear promising, at least for applications in which only 70% SO_2 removal is required, much development work remains to be completed.

Most spray dryer designs for FGD are direct adaptations of the standard designs so widely used in other industries (4). Typically in these uses, a hot gas passes downward through a cylindrical vessel, mixing with a solution or slurry atomized by rotary atomizers or nozzles. The liquid is evaporated while the droplets are in suspension and the particles are collected in a conical bottom, in external collection equipment, or both. Complete evaporation in suspension is important and is achieved by suitable design and control of operating conditions. In FGD applications the latitude of these controls is limited. The flue gas temperature is fixed by boiler efficiency requirements, and may vary as the boiler load changes; the absorbent rate is controlled by SO_x removal requirements. In addition, it is economically important that absorbent consumption be minimized. These limitations may complicate applications in which high SO_x removal efficiencies are required.

The reactions of SO_x and HCl with the absorbent proceed rapidly while surface liquid is present, but more slowly when the absorbent is dry. An important design consideration is, therefore, that the saturation temperature be approached as closely as possible and that the particles remain in contact with the flue gas as long as possible. Whatever these conditions, however, effective reaction rates require a reactive absorbent. Limestone has not proven satisfactory, leaving soda ash and lime as the only economically practical absorbents generally available in sufficient quantities. Soda ash is more reactive but also more expensive and the soluble waste of sodium sulfites and sulfates produced has raised questions of its practicality for disposal in areas of high rainfall. Lime (CaO) is less reactive and more difficult to handle because it must be slaked and then handled as a slurry. It is, however, less expensive and it produces a relatively insoluble waste of calcium sulfites and sulfates. At higher SO_x removal efficiencies the absorbent must sometimes be used at high stoichiometric ratios, particularly if lime is used. Utilization can sometimes be increased by reslurrying and recycling the waste. Also, if a highly alkaline fly ash is produced, as is usual with western coals, the fly ash alkalinity can supplement the absorbent. Coal moisture content is also a factor. Flue gas produced by lower-rank, high-moisture coals has a higher saturation temperature, limiting the amount of water that can be added, compared with high-rank coals. This may place restrictions on absorbent liquid concentrations that affect the SO_2 removal efficiency.

The methods by which vendors treat these considerations differ. In most cases, a conventional spray dryer design, rotary atomizers, lime absorbent, and fabric filter baghouse particulate collection are used. The approach to saturation is controlled by controlling water addition rates. Some warm (300°F) flue gas from the air heater may be bypassed around the spray dryers and recombined with the cleaned gas for reheating before the flue gas enters the baghouse. In extreme cases, for high SO_x removal efficiencies some hot (700°F) flue gas may be bypassed around the air heater (at the expense of boiler thermal efficiency) and the spray dryers to attain sufficient reheating.

Important exceptions to the above design exist. One important vendor uses two fluid nozzles and ESP collection and one commercial unit will use soda ash. Absorbent utilization is increased in some cases by recycling the waste, while in other cases it is not. The degree to which saturation temperature is approached (and thus the possibility of wet upsets) varies with vendors. In general, the vendor's approach to design reflects his experience and the requirements of the particular application. The technical and economic advantages of such design variations as type of atomizer, waste recycle, and ESP or baghouse collection remain, in large part, to be demonstrated in further investigation and application.

In general, based on current practice and trends, soda ash processes will not use waste recycle, will rarely use hot (700°F) gas bypass, and will use warm gas bypass only for very high SO_x removal efficiencies under unusual flue gas conditions. Lime processes will use waste recycle for low (70%) SO_x removal requirements, will usually use warm gas bypass, and will occasionally use hot gas bypass for high (over 85%) SO_x removal efficiencies.

DEVELOPMENT AND CURRENT STATUS

The first concerted spray dryer FGD studies in the United States were begun in the early 1970's by Rockwell International (5). These, however, were part of a regenerable process study rather than the waste-producing nonregenerable processes that are the subject of this paper. It was not until 1977 that extensive spray dryer FGD studies began. These received considerable impetus when dry injection studies at the Basin Electric Power Cooperative's Leland Olds Station were expanded to include spray dryers (6). Four companies subsequently operated spray dryer FGD pilot plants there and at other Basin Electric power plants as a bid qualification requirement for the FGD units on new Basin Electric construction.

Subsequently, other organizations became active in spray dryer FGD studies. In mid-1980 ten companies or consortia (shown in Table 1) were active in spray dryer FGD investigations and six had contracted for units for nine utility and three industrial installations. These contract awards are shown in Table 2. All of the commercial utility applications are for low-sulfur western coal, as has been the preponderance of pilot-scale studies.

TABLE 1. COMPANIES ACTIVE IN SPRAY DRYER-BASED FGD SYSTEMS FOR UTILITY APPLICATIONS

Babcock & Wilcox
Buell-Envirotech/Anhydro, Inc.
Carborundum Environmental Systems
Combustion Engineering, Inc.
Ecolaire Environmental Corporation
Flakt, Inc.
Joy Manufacturing/Niro Atomizer, Inc.
Research-Cottrell
Rockwell International
Wheelabrator-Frye, Inc.

TABLE 2. CONTRACT AWARDS FOR SPRAY DRYER-BASED FGD SYSTEMS

Installation	Size, gross MW	Fuel type (% S)	SO_2 removal, %	Alkali raw material	Startup date	Vendor
Utility Boiler						
Coyote Unit 1	410	Lignite (0.78)	50	Soda ash	4/81	RI/WF[a]
Antelope Valley Unit 1	440	Lignite (0.68)	62	Lime	4/82	Joy/Niro[b]
Laramie River Unit 3	575	Subbituminous (0.54)	85	Lime	4/81	B&W[c]
Stanton Unit 2	63	Lignite (0.77)	73	Lime	9/82	R-C[d]
Springerville Unit 1	350	Subbituminous (0.69)	61	Lime	2/85	Joy/Niro
Springerville Unit 2	350	Subbituminous (0.69)	61	Lime	9/86	Joy/Niro
Rawhide Unit 1	260	Subbituminous (0.29)	80	Lime	12/83	Joy/Niro
Craig Unit 3	600	Bituminous (0.70)	87	Lime	4/83	B&W
Holcomb Unit 1	280	Subbituminous (0.30)	80	Lime	6/83	Joy/Niro
Industrial Boiler						
Strathmore Paper Co.	14[e]	Bituminous (2.0-2.5)	75	Lime	7/79	Mikropul
Celanese	22[e]	Bituminous (1.0-2.0)	70-80	Lime	1/80	RI/WF
University of Minnesota	83[e]	Subbituminous (0.6-0.7)	70	Lime	9/81	Carborundum

Based on contact with vendors representing the status of announced contracts through June 1980.

a. Rockwell International/Wheelabrator-Frye.
b. Western Precipitation Division of Joy Manufacturing Company/Niro Atomizer, Inc.
c. Babcock & Wilcox.
d. Research-Cottrell.
e. Based on 2,900 aft^3/MW.

Of the contracted utility units, all use lime except the installation at the Coyote Station at Beulah, North Dakota. All except Babcock & Wilcox use atomizers in spray dryers of conventional design and fabric filter baghouse collection. B&W uses two-fluid nozzles, evolved from boiler oil burners, in horizontal chambers and either ESP's or baghouses for particulate collection. B&W manufactures all of its FGD equipment. The other vendors are either consortia which include a spray dryer manufacturer or they have an exclusive agreement with a spray dryer manufacturer.

B&W began spray dryer FGD studies in 1977, initially with a commercial spray dryer/reactor and subsequently with their own two-fluid nozzle atomizer and horizontal reactor design they call a dry scrubbing reactor. Steam was first used as the atomizing fluid; more recently air has been favored. B&W uses ESP's for particulate collection, believing them to be a more developed technology and more amenable to wet upset. B&W has not generally favored waste recycle. Rather they design for a closer approach to saturation temperature than most vendors and emphasize particle size reduction to attain efficient absorbent utilization. In addition to their two pilot units, B&W is conducting continuing studies at their Alliance, Ohio, research center. B&W has been awarded contracts for two utility applications.

The Buell Emission Control Division of Envirotech Corporation and Anhydro, Inc., of Copenhagen, Denmark, are currently developing a spray dryer FGD system as a joint venture. Buell is a designer and marketer of particulate control equipment while Anhydro is a designer and marketer of spray dryers. The pilot unit at the Martin Drake Station uses a single rotary atomizer and baghouse particulate collection.

Carborundum Environmental Systems is a subsidiary of Kennecott Copper Corporation based in Knoxville, Tennessee. Carborundum has recently signed a licensing agreement with Kochiwa Kakohki Company, Inc., a Japanese spray dryer manufacturer. The spray dryers for the Carborundum system will be manufactured in the United States while the rotary atomizers may be manufactured in either Japan or the United States. Baghouses for the spray dryer FGD system will be designed and built by Carborundum. Much of the development work for Carborundum's spray dryer FGD system was done at a 100 ft^3/min bench-scale unit at their test facility at the University of Tennessee in Knoxville. The initial pilot studies were made to qualify for bids on Basin Electric units. The present design uses a conventional spray dryer with three rotary atomizers, baghouse particulate collection, and has no waste recycle. Carborundum has been awarded a contract for an industrial boiler application.

While Combustion Engineering, Inc., has been installing limestone-based FGD systems for several years, they first entered into the development of spray dryer FGD systems in 1978. Construction of their first pilot-plant unit began in February 1979. During 1979 a license agreement was concluded with James Howden Holima BV (The Netherlands) for use of their baghouse technology. The current design consists of a conventional spray dryer with multiple nozzle atomizers and baghouse particulate collection. Compressed air is the atomizing fluid. A 30-MW demonstration unit is planned.

Ecolaire Environmental Corporation is a subsidiary of Ecolaire, Inc., which markets Ecolaire's spray dryer FGD process. Other subsidiaries in the Ecolaire Corporation have been supplying equipment to the electric utility industry for many years. Before the design and construction of their mobile demonstration unit (MDU) in 1979, Ecolaire had very little experience in the design and operation of FGD systems. The MDU was erected in 1979 at a Nebraska power plant. The unit has a conventional spray dryer using either a rotary or two-fluid nozzle atomizer and fabric filters for particulate collection.

The Western Precipitation Division of The Joy Manufacturing Company, which markets fabric filter baghouses, and Niro Atomizer, Inc., which markets spray dryers, have an exclusive agreement to market a spray dryer FGD system. Niro began FGD studies in Denmark in 1975. The first pilot plant was operated in 1978 to qualify to bid for Basin Electric units. The Joy/Niro design consists of a spray dryer of conventional design using one rotary atomizer and a manifold that introduces flue gas above and below the atomizer. Part of the particulate matter is collected in the bottom of the spray dryer and the rest is collected in baghouses. Waste recycle, using the large particles from the bottom of the spray dryer, is used for most applications. Current development work is being conducted in Niro's Copenhagen laboratory. A demonstration unit at the Riverside Power Station will provide the facilities for future large-scale testing. Joy/Niro has been awarded five commercial contracts for utility systems.

The Research-Cottrell system uses spray dryers of conventional Komline-Sanderson design with either a single or multiple rotary atomizers. Part of the particulate matter is collected in the bottom of the spray dryer and the rest is collected in baghouses. Waste recycle is usually used. Most details and test results are proprietary and little published information is available. The pilot plant at the Comanche Station is partially funded by EPA, so results of these tests will probably be available in the near future. Research-Cottrell has been awarded one contract for a utility boiler.

Until early 1980 Rockwell International and Wheelabrator Frye, Inc., had agreements to market spray dryer FGD systems based on Rockwell International's experience in spray dryer FGD and Wheelabrator Frye's fabric filter technology. This joint venture was dissolved in 1980 and each will now market its own system. The first RI/WF spray dryer pilot unit was operated at the Leland Olds Station in 1977. This and other pilot studies have provided considerable data on the system. The design consists of a conventional spray dryer with three rotary atomizers and baghouse particulate collection. Waste recycle is used if the conditions warrant it. RI/WF have been awarded two commercial contracts, one for utility system and one for an industrial application.

DESIGN AND ECONOMIC PREMISES FOR THE FGD EVALUATION

The economic evaluations are based on flue gas cleaning (FGD and fly ash) systems to meet the 1979 NSPS for a new 500-MW pulverized-coal-fired, dry-bottom utility boiler. The FGD systems are designed with one redundant train, 50% emergency flue gas bypass, and are costed as proven technology with no adjustments for estimated stage of development. The power plant is assumed to have a 30-year, 165,000-hour life and to operate 5,500 hours the first year. Flue gas compositions are based on a 0.7% sulfur, 9.7% ash, 9,700 Btu/lb western coal; a 0.7% sulfur, 16% ash, 11,700 Btu/lb eastern coal; and a 3.5% sulfur, 16% ash, 11,700 Btu/lb eastern coal. A Northern Great Plains location is used for the western coal case; a midwestern location is used for the eastern coal cases. The spray dryer designs are generic and are based on vendor information. The limestone scrubbing process is based on data from the EPA Shawnee test facility and general industry information. Design data for the absorbers are shown in Table 3.

Raw materials consist of commercial-grade soda ash at $145/ton (1984 dollars), pebble lime at $102/ton in the West and $75/ton in the East; and limestone at $8.50/ton. The waste disposal sites are one mile from the FGD facility. They consist of a clay-lined pond for the soda ash spray dryer process and landfills for the lime spray dryer and limestone scrubbing wastes.

The economics consist of study-grade capital investments, first-year annual revenue requirements, and levelized annual revenue requirements. The capital investments are based on major equipment costs developed from flow diagrams and material balances and factored costs for installation and ancillary equipment. The capital investments are estimated to have an absolute accuracy of -20% to +40%. However, since the same estimation methods are used for each evaluation, the accuracy for comparison is probably much better, i.e., ±10%. Capital investment costs are based on mid-1982 costs.

First-year annual revenue requirements consist of raw material, operating, and overhead costs and levelized capital charges. The levelized annual revenue requirements are factored to account for a 10% discount and a 6% inflation rate over the life of the power plant.

SYSTEMS ESTIMATED

For the low-sulfur western coal case, the soda ash and lime spray dryer processes and the limestone scrubbing process are evaluated. For the eastern coal cases only the lime spray dryer and limestone scrubbing processes are evaluated because of the economically indefinable problems likely to occur with soluble sodium wastes in high rainfall regions. Process designs include all the equipment and material needed to transfer the flue gas from the boiler wall to the stack plenum. All requirements for both fly ash collection and disposal and SO_2 removal and disposal are included in the costs.

TABLE 3. FGD SYSTEM DESIGN CONDITIONS

	Low-sulfur western coal			Low-sulfur eastern coal		High-sulfur eastern coal	
	Soda ash spray dryer	Lime spray dryer	Limestone scrubbing	Lime spray dryer	Limestone scrubbing	Lime spray dryer	Limestone scrubbing
Absorbent stoichiometry	1.0	1.2	1.1	1.3	1.1	1.5	1.3
Bypass, %	0	22	27.9	19	25.2	4[a]	0
Total FGD ΔP, in. H_2O	12	12	6.5	12	6.5	12	6.5
Absorber							
Removal efficiency, %	70	83	90	83	90	89	90
Absorbent liquid, % solids[b]	0	22.5	60	3	60	17	60
ΔP, in. H_2O	2	2	2	2	2	2	2
L/G, gal/kaft^3[b]	0.13	0.2	80	0.3	80	0.3	80
Exit gas, wt % liquid	0	0	0.1	0	0.1	0	0.1
Effluent, % solids	100	100	15	100	15	100	15
Recombined gas, °F[c]	170	170	170	170	160	170	127
Reheat, °F	-	-	-	-	10	-	43

a. Hot gas bypass.
b. Excludes dilution water and recycle loop if used.
c. The ID fan provides an additional 5°F of reheat to achieve a stack temperature of 175°F.

The soda ash and lime spray dryer processes consist of four or five trains of cylindrical spray dryers, each with three rotary atomizers. One train is included as a nonoperating spare in all cases. Hot gas (700°F) bypass of flue gas around the air heater and spray dryers is designed into the low-sulfur coal cases although under normal operating conditions it will not be needed. A continuous 4% hot gas bypass is used in the lime spray dryer process for the high-sulfur eastern coal case. Warm gas (300°F) bypass rates of 22% and 19% are used in the lime spray dryer process for the low-sulfur western and eastern coal cases respectively. No flue gas bypass is used for the soda ash process under normal operating conditions. The high reactivity of the soda ash does not require a close approach to the saturation temperature and flue gas bypass is not necessary to ensure dry conditions in the baghouse. In all cases the processes are designed for a flue gas stack temperature of 175°F. A single baghouse is used for particulate collection. The waste is pneumatically conveyed to storage silos and trucked to a disposal site one mile away. An earthen-diked, clay-lined pond is used for the soda ash spray dryer process and a landfill is used for the lime spray dryer process. For the lime spray dryer process in the low-sulfur western coal case, waste recycle is used to reduce absorbent consumption. Waste is not recycled in the other processes.

The limestone scrubbing process consists of four or five trains of spray tower absorbers, one of which is a spare, using a 15% solids slurry of ground limestone as the reactant. The sulfite in the slurry is oxidized by sparging air into the circulation tank to produce $CaSO_4 \cdot 2H_2O$. A purge stream is dewatered by thickening and filtering to 80% solids. The waste is trucked one mile to a landfill and disposed of, along with fly ash collected in ESP upstream of the FGD system. For the low-sulfur western and eastern coal cases, 28% and 25% warm gas bypass is used, respectively, and the scrubbing efficiency is 90% for both cases to obtain an overall 70% SO_2 reduction. Full scrubbing at an 89% SO_2 reduction efficiency is used for the high-sulfur eastern coal case. The flue gas bypass eliminates the need for flue gas reheat in the low-sulfur western coal case and substantially reduces it in the low-sulfur eastern coal case. Full reheat is used in the high-sulfur eastern coal case.

ECONOMIC EVALUATION

Capital investments, first-year annual revenue requirements, and levelized annual revenue requirements were developed based on the processes described in the systems estimated section and the conditions described in the premises. (Due to the preliminary nature of the study, the results in this paper are given as ranges.)

Capital Investment

The capital investments for the soda ash and lime spray dryer processes and the limestone scrubbing process are shown in Table 4.

TABLE 4. CAPITAL INVESTMENT

	Capital investment	
	M$	$/kW
Low-Sulfur Western Coal		
Soda ash spray dryer	76-80	152-160
Lime spray dryer	72-76	144-152
Limestone scrubbing	84-88	168-176
Low-Sulfur Eastern Coal		
Lime spray dryer	72-76	144-152
Limestone scrubbing	90-94	180-188
High-Sulfur Eastern Coal		
Lime spray dryer	90-94	180-188
Limestone scrubbing	118-122	236-244

Note: Ranges are used to quantify the preliminary results given in this paper.

In comparing the soda ash and lime spray dryer processes for the low-sulfur western coal case, the overall capital investments differ only slightly. The largest cost area, particulate collection, is the same for both. Similarly, the total of the gas handling and SO_2 absorption areas differ little, suggesting little cost difference between partial bypass and full scrubbing. Feed preparation and handling costs are lower for the soda ash spray dryer process but this is more than offset by the higher disposal site construction and land costs because a pond is used for the sodium wastes. Waste recycle in the lime process approximately doubles the solid handling costs but costs in this area are small compared with other areas. The comparisons suggest that neither bypass nor waste recycle is an important capital investment consideration.

In comparisons of the lime spray dryer and limestone processes, more significant differences emerge. In overall capital investment the limestone scrubbing process is 16% to 31% higher than the lime spray dryer process, the difference increasing as the coal sulfur content increases. The major cost area for the limestone scrubbing process is SO_2 absorption, representing over one-third of the direct costs. In addition, this cost increases about one-half as the coal sulfur content increases from 0.7% to 3.5%. In contrast, the SO_2 absorption area costs for the lime spray dryer process are about one-half those of the limestone scrubbing process and they increase little with coal sulfur content. These SO_2 absorption area costs are the major cause of the capital investment cost differences between the processes.

In other areas, the two processes have similar costs. The limestone scrubbing process has moderately higher gas handling costs, very slightly higher costs for wet solids separation (thickening and filtering) compared with dry solids handling (pneumatic conveying and silo storage) and slightly lower disposal costs because of the higher bulk density of the gypsum waste from the limestone process. Materials handling costs for the limestone scrubbing process are lower because the limestone can be simply stockpiled. Limestone grinding costs greatly exceed lime slaking costs (due to the inclusion of a redundant feed preparation area in the limestone scrubbing process). However, the sum of the costs for handling and preparing absorbents are similar.

Annual Revenue Requirements

First-year and levelized annual revenue requirements are shown in Table 5. Levelized costs are first-year costs adjusted by a factor of 1.886 times direct operating expenses, an adjustment that takes into account 6% per year inflation and discounting by 10% per year (cost of money) over the 30-year life of the installation.

In the low-sulfur western coal case, the lime spray dryer process has the lowest first-year annual revenue requirements, followed by the soda ash spray dryer and limestone scrubbing processes respectively. For the spray dryer processes the difference is almost entirely the result of absorbent costs (1 mill/kWh for soda ash and 0.4 mill/kWh for lime). Other minor differences account for the remaining cost difference. For the limestone scrubbing process, absorbent costs are minor, less than 0.1 mill/kWh, but maintenance costs are, predicted to be, more than double those of the spray dryer processes. These, along with the indirect costs and the capital charges, account for the cost differences between the limestone scrubbing process and the spray dryer processes.

For the low-sulfur eastern coal case a similar relationship prevails. Most costs differ insignificantly from those of the low-sulfur western coal case, in spite of the different flue gas quantities. The lime spray dryer costs are slightly lower, primarily because of the lower lime cost in the East. The limestone scrubbing process costs are slightly higher, a result of general cost increases stemming from the lower flue gas bypass. The small amount of flue gas reheat required for the limestone scrubbing process does not significantly affect process costs.

TABLE 5. ANNUAL REVENUE REQUIREMENTS

	Annual revenue requirements			
	First year		Levelized	
	M$	Mills/kWh	M$	Mills/kWh
Low-Sulfur Western Coal				
Soda ash spray dryer	19-20	6.9-7.1	26-27	9.4-9.8
Lime spray dryer	18-19	6.5-6.9	24-25	8.7-9.1
Limestone scrubbing	21-22	7.6-8.0	29-30	10.5-10.9
Low-Sulfur Eastern Coal				
Lime spray dryer	17-18	6.1-6.5	23-24	8.4-8.7
Limestone scrubbing	23-24	8.4-8.7	31-32	11.3-11.6
High-Sulfur Eastern Coal				
Lime spray dryer	27-28	9.8-10.2	40-41	14.5-14.9
Limestone scrubbing	32-33	11.6-12.0	45-46	16.4-16.7

Note: Ranges are used to quantify the preliminary results given in this paper.

For the high-sulfur eastern coal case, the cost advantage of the lime spray dryer process, compared with the limestone scrubbing process, is decreased. The increase in cost for both the lime spray dryer process and the limestone scrubbing process in going from the low-sulfur eastern coal case to the high-sulfur eastern coal case is about 50%. The salient cost factor is absorbent cost. Absorbent costs for both processes increase about seven times. For the lime spray dryer process, however, this increase results in absorbent costs totaling 24% of the total annual revenue requirements; for the limestone scrubbing process only 3%. Other costs increase little in comparison and in general the increases are similar for both processes. A significant requirement for flue gas reheat also appears in both processes, one in the form of steam, the other in the form of hot flue gas.

CONCLUSIONS

The development of spray dryer FGD has been rapid and processes by several vendors will soon be in operation. The technical and economic feasibility of the vendors' approach to design features such as type of atomizer, degree of approach to saturation temperature, the particulate collection method, and waste recycle remain to be demonstrated. Current trends suggest the majority will use a lime slurry with rotary atomizers, partial flue gas bypass, and fabric filter collection. In addition, little data on high-sulfur coal applications are available. Spray dryer FGD will probably be limited largely to low-sulfur coal applications until better data are developed.

Interest in dry injection continues but development of these processes has been slow. Nahcolite, the most promising candidate absorbent, is unlikely to be available in sufficient quantities for several years. The development of processes using other absorbents, including limestone, is only beginning and the practicality of such processes remains to be proved.

The spray dryer processes are similar in cost and both are substantially lower in capital investment and annual revenue requirements than the limestone scrubbing process, at least for low-sulfur applications. The lime spray dryer process is more cost effective than the limestone slurry process for both the low- and high-sulfur eastern coal case. This latter conclusion is an unexpected result based on discussions with process vendors. The differences are largely the result of lower spray dryer equipment costs, compared with wet scrubbers, and lower utility and maintenance costs. Offsetting these advantages, absorbent costs become substantial for the spray dryer processes at high sulfur coal levels. The relationship of equipment costs is unlikely to change substantially. Operating and absorbent costs could, however, require adjustment as more operating experience is gained.

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SPRAY DRYER FGD CAPITAL AND OPERATING COST ESTIMATES FOR A NORTHEASTERN UTILITY

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ABSTRACT

This paper is in response to a request by the EPA Office of Environmental Engineering and Technology (OEET) to develop investment and operating costs for a spray dryer flue gas desulfurization (FGD) system retrofitted on several boilers in two power stations of a northeastern utility. These power stations are among 23 power plants involved in DOE proposed coal conversion actions under the Fuel Use Act of 1978. Based on budget estimates received from a number of FGD system vendors, total installed cost of this system (which would control SO_x and particulate emissions to present uncontrolled levels produced by No. 6 oil-firing), would range from \$153.9 million (\$89/Kw) to \$204.0 million (\$118/Kw) for a generalized northeast location, with a significantly higher cost of \$295.8 million (\$171/Kw) for a highly urbanized location typified by the northern New Jersey - southern New York area. Based on projected oil costs, coal costs, plant renovation costs, and annualized operating costs of the emission controls, potential savings by coal conversion would be \$187 million annually (19 mills/Kw-hr.) Problems of FGD waste disposal as well as site-specific problems are also considered in this paper.

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SPRAY DRYER FGD CAPITAL AND OPERATING COST ESTIMATES FOR A NORTHEASTERN UTILITY

INTRODUCTION

There are 23 power plants in the northeastern part of the United States that are subject to the Department of Energy (DOE) proposed prohibition orders issued under the Power Plants and Industrial Fuel Use Act of 1978 (FUA). These 23 power plants are also included in Phase 1 of the proposed Power Plant Petroleum and Natural Gas Displacement Act of 1980 (also known as the utility oil and gas backout bill). The intent of the DOE proposed prohibition orders and the proposed oil and gas backout legislation is to facilitate the conversion of these coal-convertible boilers from oil or gas to coal as the primary energy source.

The DOE proposed prohibition orders and pending oil backout legislation directly involve EPA, state, and local regulations covering SO_2 and particulate emission limitations. The utilities affected will be required to identify the available SO_2 and particulate control technologies that would be the "best available control technology" (BACT) for each specific utility power plant and to retrofit an appropriate technology accordingly. Among the utilities concerned is a northeastern utility located in the highly urbanized northern New Jersey - southern New York area with two power stations under the DOE proposed prohibition orders. This utility requested that EPA Region II provide assistance in the development of costs and other information on the spray dryer flue gas desulfurization (FGD) process to be used in connection with the proposed conversion of these two stations from the use of low-sulfur No. 6 fuel oil as fuel (oil-firing) to the use of low-sulfur coal as fuel (coal-firing). EPA Region II referred this request to the EPA Office of Environmental Engineering and Technology (OEET) with the proviso that the name of the utility involved be kept confidential from potential vendors of the spray dryer FGD system and that vendor names be kept confidential from the utility.

The MITRE Corporation, in conjunction with OEET, developed the requested information. The basis of this effort involved the preparation of a set of bid specifications covering the proposed addition to the utility of spray dryer FGD systems to treat the flue gas from a total of four boilers at the two power stations under consideration. These specifications were submitted to a total of five vendors, four of which responded with budget estimates covering the direct capital cost of spray dryer FGD systems (1980 dollars) and with estimates of annual operation and maintenance (O&M) costs (1982 dollars).

In addition to the development of spray dryer FGD system costs, MITRE discussed the site-specific issues in the proposed conversion with the utility involved. The basic technical and economic premises employed in the MITRE study (other than those supplied by the utility) were reviewed with the utility as to their accuracy and applicability to the specific boiler sites.

The succeeding sections of this paper present the following information:

- Technical and economic premises used in the development of the presented costs
- Installed and annualized operating costs determined for the utility boilers involved
- Analysis of the comparative costs of oil-firing versus low-sulfur coal-firing used in conjunction with spray dryer FGD
- Presentation of the problems involved in retrofitting the spray dryer FGD process at the particular sites
- Conclusions reached in the study

TECHNICAL AND ECONOMIC PREMISES USED IN DEVELOPING SPRAY DRYER FGD COSTS FOR A NORTHEASTERN UTILITY

This section presents the basis used in a cost analysis of a spray dryer FGD system applicable to a large northeastern utility which is considering the conversion from oil to low-sulfur coal firing. The development of capital and O&M cost ranges is based on the capability of the spray dryer FGD system to control to the present levels of uncontrolled SO_2 and particulate emissions from the two utility power stations that are currently using low-sulfur No. 6 fuel oil as the boiler fuel. The technical and economic premises used in this analysis are presented below.

Coal data and information on plant design parameters pertinent to the development of spray dryer FGD cost estimates were supplied by the northeastern utility for the two power stations involved. These data are tabulated in Tables 1, 2 and 3.

TABLE 1

COAL AND ASH CHARACTERISTICS (AS FIRED) OF PROPOSED LOW SULFUR COAL

(Coal Source: West Virginia)

Proximate Analysis/Percent by Weight:

	Typical	Range
Fixed Carbon	51.94	49.00 - 54.50
Volatile Matter	30.56	24.00 - 34.00
Moisture	5.00	2.00 - 10.00
Ash	12.50	8.00 - 18.00
Heating Value, Btu/lb	12,200	11,500 - 14,500
Sulfur, Percent by Wt	0.90	0.60 - 1.00

Ultimate Analysis/Percent by Weight:

	Typical	Range
Carbon	69.74	65.00 - 75.00
Moisture	5.00	2.00 - 10.00
Hydrogen	4.40	4.00 - 6.00
Oxygen	6.02	4.00 - 12.00
Nitrogen	1.33	1.10 - 1.50
Chlorine	0.11	0.07 - 0.13
Sulfur	0.90	0.60 - 1.00
Ash	12.50	8.00 - 18.00
Total	100.00	

Ash Mineral Analysis/Weight Percent (Ignited Basis):

Phosphorus Pentoxide, P_2O_5	0.30
Silica, SiO_2	52.96
Ferric Oxide, Fe_2O_3	4.66
Alumina, AL_2O_3	35.22
Titania, TiO_2	2.01
Lime, CaO	1.02
Magnesia, MgO	0.35
Sulfur Trioxide, SO_3	1.57
Potassium Oxide, K_2O	0.77
Sodium Oxide, Na_2O	0.69
Undetermined	0.45
Total	100.00

TABLE 2

BOILER PERFORMANCE DATA

Parameter	Power Plant A Unit #1	Unit #2	Power Plant B
Boiler Size, Mw (net sustained rating)	335	491	900
Design Efficientcy, %[a]	86.5	89.5	90
Station Efficiency (overall conversion to electricity), %[a]	34.1		34.3
Station Heat Rate, Btu/kwh[b]	10,264		10,265
Percent excess air	25		25
Max. Fuel Firing Rate, lb/hr	678,250(combined)		821,200
Anticipated Load Factor, percent (based on coal-firing)	65	65	65
Years of Operating Life Remaining	24	34	30

[a] 1978 data.
[b] 1977 data.

TABLE 3

BOILER FLUE GAS DATA

Parameter	Power Plant A	Power Plant B
Max. Stack Temp., °F	270	294
Average Stack Temp., °F	265	290
Minimum Stack Temp., °F	230	245
Max. Gas Flow, 10^6 acfm[a]	2.46[b]	2.9
Overall pressure drop, inches H_2O	20	20
Required particulate emission limit, lb/10^6 BTU[c]	0.04	0.03
Percent particulate reduction rqd.	99.5	99.5
Required SO_2 emission limit, lb/10^6 BTU[d]	0.3	0.3
Percent SO_2 reduction required	82	82

[a] Corresponding to maximum stack temperature.
[b] Both units discharge to common stack.
[c] These are the utilities present performance data; the legal limit is 0.1 lb/10^6 BTU.
[d] Corresponding to the present SO_2 emissions from burning of low-sulfur No. 6 fuel oil.

The spray dryer FGD process is provided with system redundancy: two of the designs submitted have provisions for an extra spray dryer module; while the remaining two designs have provisions for an extra atomizer head in each of the spray dryer modules. Provisions for turndown and maintenance are limited to inclusion of a common plenum between systems with dampers to allow for the shut down of individual spray dryer for maintenance. The fabric filter designs also allow for the shut down of individual compartments for maintenance.

Economic Premises

The spray dryer FGD process vendors contacted during the course of this study were asked to provide budget estimates (in 1980 dollars) with a range of accuracy of ± 30 percent. The vendors were also asked to provide data with which to estimated direct O&M costs projected to 1982 dollars. The economic assumption (other than those used in vendor calculations) used to calculate annualized operating costs are tabulated in Table 4.

Table 5 presents the TVA methodology used in deriving spray dryer FGD system total installed costs from direct capital costs supplied by vendors. The overall ratio of total installed cost to direct capital cost in the TVA methodology is 2.0. This value was used in the present paper in all cases to adjust vendor costs, except where vendors indicated that some indirect cost items (as listed in Table 5) are included in their direct capital cost

TABLE 4

ECONOMIC ASSUMPTIONS USED IN ESTIMATING ANNUALIZED OPERATING COSTS (1982 DOLLARS)

Item[a]	Value
Pebble Lime Sorbent at 90% purity ($/ton)	61
Ca/S Mole Ratio in Spray Dryer Feed	~1.0
Labor Rates ($/hr.)	
Plant Operators	10
Supervisors	15
Analyses	10
Power Cost ($/Kw-hr.)	0.10
Process Water Cost ($/1000 gallons)	0.25
Overhead Factor (% of labor and maintenance costs)	60
Levelized capital charge factor	0.147[b]

[a] These are in addition to assumptions made by the various vendors in developing O&M cost estimates.

[b] Derived from TVA economic model for regulated utility (EPA-600/7-80-050). This factor includes charges for the capital recovery factor (based on a publicly-owned utility), interim equipment replacements, insurance and property taxes, State and Federal income taxes, and credits for investment credits and accelerated depreciation.

estimates, e.g. engineering design and supervision. In these cases, the ratio of total installed cost to direct cost was adjusted based on the percentages shown in Table 5.

INVESTMENT AND ANNUALIZED OPERATING COSTS OF A SPRAY DRYER FGD SYSTEM AS APPLIED TO THE PROPOSED CONVERSION FROM OIL TO COAL-FIRING FOR A NORTHEASTERN UTILITY

The technical information presented in Tables 1 through 3 was originally submitted to four vendors of spray dryer FGD systems. These vendors were asked to supply estimates for a spray dryer system to be located in the northeastern region, i.e., to provide estimates for a system that includes spray dryers for SO_2 removal and a fabric filter to collect both the solid calcium sulfite/sulfate waste produced in the spray dryer and the fly ash originally present in the flue gas from the boilers. The estimates were to be in 1980 dollars and of budget estimate accuracy, i.e., within ± 30 percent of the projected final cost of the battery-limits facility. With respect to operating costs, the vendors were asked to supply estimates of appropriate utilities, labor and maintenance requirements costs. If O&M costs were supplied, these were to be projected to 1982 dollars. Finally, the vendors were asked to supply estimates of the overall area required for a normal spray dryer FGD installation based on the assumption that adequate vacant land is available at the utility power stations.

Spray Dryer FGD System Total Installed Cost Estimates

Three of the four vendors originally contacted, submitted the capital cost information requested. The direct capital costs quoted

TABLE 5

INDIRECT COST ITEMS INCLUDED IN SPRAY DRYER FGD TOTAL INSTALLATION COST ESTIMATES[a]

Cost Item No.	Item Description	Value Used in TVA Cost Estimates	Percent of Total Indirect Cost
1	Engineering Design and Supervision	7% of direct investment	6.8
2	Architect and Engineering Contractor	2% of direct investment	1.9
3	Construction Expense	16% of direct investment	15.4
4	Contractor Fees	5% of direct investment	4.8
5	Contingency	20% of direct investment	25.1
6	Startup and Modification Allowance	10% of total of direct investment plus items 1 through 5	15.0
7	Interest During Construction	15.6% of total of direct investment plus items 1 though 5	23.6
8	Royalties	0.5% of direct investment	0.4
9	Land	$5,000/acre	1.5
10	Working Capital	Equivalent cost of 1 months raw material, 1.5 months labor and utilities, and 1.5 months plant and administrative costs	5.5
			100.0
	Ratio of Indirect Costs (Items 1 through 10) to Direct Capital Cost		1.0
	Ratio of Total Installed Cost to Direct Capital Cost		2.0

[a]Source: "Preliminary Economic Analysis of a Lime Spray Dryer FGD System", T.D. Burnett and W.E. O'Brien, EPA-600/7-80-050, March, 1980.

were for a spray dryer FGD system that begins at the inlet flange of the plenum feeding the spray dryers with boiler flue gas and extends to the outlet flange of the fabric filter. The budget estimates generally included all major pieces of equipment, installation labor, process controls, instrumentation, electrical, interconnecting ductwork and piping, painting, enclosures, and insulation. The estimates did not include the ID fan (located at the outlet of the fabric filter), the stack, foundations and site preparation, electrical supply, utility supply and return headers, lime delivery and unloading systems, and FGD waste disposal systems. Tables 6 and 7 present a summary of the major equipment items in the spray dryer FGD systems proposed by the three vendors (referred to in this paper as Vendors No. 1, 2 and 3). Table 8 presents the estimated direct capital costs supplied by these vendors and the estimated total installed costs developed in this paper through the use of TVA estimates of indirect costs that are not normally included in vendors estimates (see Table 5).

A set of investment cost estimates (in parenthesis in Table 8) are included for Vendor No. 3 and for a fourth vendor, (Vendor No. 4). These estimates were obtained after it was determined from conversations with the northeastern utility that the unique labor cost/productivity situation related to plant construction in the densely urbanized area involved would cause exceptionally high total construction costs for the proposed spray dryer FGD facilities

TABLE 6

ANALYSIS OF VENDOR QUOTATIONS: BUDGET ESTIMATES OF SPRAY DRYER FGD SYSTEMS FOR A NORTHEASTERN UTILITY

SUMMARY OF CAPITAL EQUIPMENT INCLUDED IN ESTIMATES

Capital Equipment Item	Vendor No. 1			Vendor No. 2	
	Plant A		Plant B	Plant A	Plant B
	Unit No. 1	Unit No. 2		Units No. 1 & 2[a]	
Spray Dryers	4 plus 1 spare	5 plus 1 spare	9 plus 1 spare	6-46 foot diameter each equipped with 3 atomizers plus 1 spare atomizer	6-46 foot diameter each equipped with 3 atomizers plus 1 spare atomizer
Fabric Filter	1	1	1	1-20 compartment	1-20 compartment
Lime Feed Preparatory System	1	1	1	2-lime slakers 2-lime slurry feed tanks 2-lime surge bins	2-lime slakers 2-lime slurry feed tanks 2-lime surge bins
Recycled Waste System	--	--	--	1-recycled waste storage silo 1-recycle slurry tank	1-recycled waste storage silo 1-recycle slurry tank
Solids Conveying System	Yes	Yes	Yes	Yes	Yes
Instrumentation and Controls	Yes	Yes	Yes	Yes	Yes
All Interconnecting Ducting and Piping	Yes	Yes	Yes	Yes	Yes
Estimated Plant Area Required, Sq. ft.	65,000	90,000	114,000	107,000	107,000

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[a] Flue gas output from utility boiler units 1 and 2 is combined and treated in a common spray dryer FGD system.

TABLE 7

ANALYSIS OF VENDOR QUOTATIONS: BUDGET ESTIMATES OF SPRAY DRYER FGD SYSTEMS FOR A NORTHEASTERN UTILITY

SUMMARY OF CAPITAL EQUIPMENT INCLUDED IN ESTIMATES

Capital Equipment Item	Vendor No. 3		Vendor No. 4	
	Plant A Units No. 1 & 2[a]	Plant B	Units No. 1 & 2[a]	Plant B
Spray Dryers	4-46 foot diameter by 36 foot cylindrical height, each equipped with one rotary atomizer. One spare unit of identical design is provided.	5-46 foot diameter by 36 foot cylindrical height, each equipped with one rotary atomizer. One spare unit of identical design is provided.	4-48 foot diameter each equipped with 3 atomizers plus 1 spare atomizer	No quotation supplied
Fabric Filter	2-24 compartment total 1.16 million square foot total	2-28 compartment total, 1.30 million square foot total	1-20 compartment, 1.14 million square feet	
Lime Feed Preparatory System	1-lime feed bin 2-lime slakers 2-lime slurry mix tanks 1-lime slurry feed tanks	1-lime feed bin 2-lime slakers 2-lime slurry mix tanks 2-lime slurry feed tanks	3-lime slakers 1-lime slurry preparation and feed tank 1-lime storage and day bin 8-slurry transfer pumps(including 4 spares)	
Recycled Waste System	1-recycled waste feed bin	1-recycled waste feed bin	1-recycled waste storage bin 1-recycle slurry tank 5-recycle slurry pumps (including 1 spare)	
Solids Conveying System	Yes	Yes	Yes	
Instrumentation and Controls	Yes	Yes	Yes	
All Interconnecting Ducting and Piping	Yes	Yes	Yes	
Estimated Plant Area Required, Sq. ft.	98,000	105,000	98,000	

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[a] Flue gas output from utility boiler units 1 and 2 is combined and treated in a common spray dryer FGD system.

TABLE 8

ANALYSIS OF FOUR VENDOR QUOTATIONS: BUDGET ESTIMATES OF SPRAY DRYER FGD SYSTEMS FOR A NORTHEASTERN UTILITY

ESTIMATES OF TOTAL CAPITAL INVESTMENT IN 1980 DOLLARS

	Vendor No. 1			Vendor No. 2		Vendor No. 3		Vendor No. 4	
	Plant A		Plant B	Plant A	Plant B	Plant A	Plant B	Plant A	Plant B
	Unit #1	Unit #2		Units #1 & 2		Units #1 & 2		Units #1 & 2	
Direct Capital Cost									
$ Millions	16.3	21.8	37.0	48.8	52.4	47.5(70.0)[c]	52.5(75.0)	(70.0)	No estimate provided
$ /Kw	49	44	41	59	58	58 (84.7)	58 (83.3)	(84.7)	
Total Installed Cost									
$ Millions[a]	33.9	44.5	75.5	84.9[b]	91.1[b]	96.9(142.8)	107.1(153.0)	(121.7)[b]	
$ /Kw	99	91	84	102.8	101.2	117 (173)	119 (170)	(147)	
Total Installed Cost (Both Plants)									
$ Millions		153.9		176.0		204.0 (295.8)			
$ /Kw		89.1		102.0		118.2 (171.4)			

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[a] Developed from the TVA cost model presented in EPA-600/7-80-050, March 1980, in which the ratio of total installed cost/direct cost is 2.04.

[b] These vendors includes 28.9% of indirect costs (based on the TVA model) in the quoted direct costs.

[c] Numbers in parenthesis are estimates for a utility site in the northern New Jersey - southern New York area.

as compared to average nationwide construction costs. Vendors No. 3 and 4 provided direct capital budget cost estimates based on this situation, and total installed costs were calculated from these values. Additionally, the Vendor No. 4 summary of spray dryer FGD system capital equipment is included in Table 7 for one of the northeastern utility's facilities (Plant A). Time limitations prevented Vendor No. 4 from performing a cost estimate for Plant B.

A review of the results presented in Table 8 indicates that spray dryer FGD process costs estimated by the first three vendors generalized northeastern regional situation show a range of total installed cost for both utility sites from $153.9 million ($89/Kw) to $204.0 million ($118/Kw). The higher estimates submittted by Vendors 3 and 4 for Plant A alone range from $121.7 million ($147/Kw) to $142.8 million ($173/Kw), and reflect the higher costs of construction in the densely urbanized locations. The total installed costs submitted by Vendor No. 3 for both plants in the urban situation are 45 to 92 percent higher than the overall range of combined total installed costs derived from estimates submitted by Vendor No. 1, 2 and 3 for spray dryer FGD facilities.

Table 9 presents actual capital and projected operating costs for three utility spray dryer FGD installations currently under construction. These systems are all being installed on western utility boilers, and are equipped to burn low sulfur (<1.0%) western lignite or sub-bituminous coal. Total installed costs

TABLE 9

SUMMARY OF INSTALLATION AND OPERATING COSTS OF UTILITY SPRAY DRYER FGD SYSTEMS CURRENTLY BEING INSTALLED[a]

SYSTEM	GENERATING CAPACITY	SORBENT	COAL	CAPITAL COST	REPORTED OPERATING COSTS
Rockwell/Wheelabrator-Frye @Otter Tail Power Co.'s Coyote Station, Unit 1, Beulah, ND Start up date: June 1981	410 MW (1,890,000 acfm)	Soda Ash	North Dakota lignite. Average S-0.78% HHV- 7050 BUT/lb. Ash-7%	43,800,000 ($107/Kw)	$6,580,000/yr (2.5 mils/kWhr)[d] Does not include waste disposal cost.
Joy/Niro @Basin Electric's Antelope Valley Station, Unit 1, Beulah, ND Start up date: April 1982.	440 MW (2,200,000 acfm)	Lime	North Dakota lignite. Average S-0.68% Maximum S-1.22%	$49,665,100 ($113/kW)[c]	$2,270,834/yr (0.8 mils/kWhr)[d] sorbent cost (lime) - $1,102,500 ($60/ton basis) Does not include waste disposal cost.
Babcock and Wilcox @Basin Electric's Laramie River Station, Unit 3, Wheatland, Wyoming. Start up date: Spring 1982.	500 MW (2,810,000 acfm)	Lime	Wyoming sub-bituminous Average S-0.54% Maximum S-0.81% HHV-8140 BTU/lb. Ash-8%	$49,807,000 ($99.6/kW)[c]	$2,571,000/yr (0.7 mils/kWhr)[d] sorbent cost (lime) - $1,396,570 ($60/ton basis) Does not include waste disposal cost.

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[a] Source: Blythe, G.M., et. al., "Survey of Dry SO_2 Control Systems", EPA-600/7-80-30, February 1980.

[b] Capital cost for complete turnkey installation from air preheater outlet to stack connection, excluding I.D. fans. Cost projected to 1982 using TVA cost index projections (EPA-600/7-80-050, p. 15)

[c] Evaluation based on 35-year life, annual plant factor of 75% (1981$). Source: Janssen and Eriksen. "Basin Electric's Involvement with Dry FGD" presented at EPA Symposium on FGD, Las Vegas, Nevada. March 1979.

[d] Operating costs in 1982 dollars.

(1981) dollars) for two spray dryers FGD systems equipped with fabric filters for particulate collection range from \$43.8 million (\$107/Kw) to \$49.7 million (\$113/Kw). The total installed cost (1981 dollars) for one system equipped with an electrostatic precipitator (ESP) for particulate collection is \$49.8 million (\$99.6/Kw). The reported costs for spray dryer-fabric filter FGD system compare closely with the estimated \$/Kw costs for the FGD systems for the generalized northeast region based on estimates submitted by Vendors No. 1, 2 and 3.

Spray Dryer FGD System Annualized Operating Cost Estimates

For the purpose of this paper, only first-year annualized operating costs were estimated, since determination of levelized costs over the life of the northeastern utility power stations involves the use of uncertain values of the time-value of money and of the inflation factor.

Annualized operating costs include direct O&M costs, capital-related costs and overhead costs. Vendors No. 1 and 3 reported 1982 projected O&M costs, while Vendors No. 2 and 4 supplied estimates of utilities and manpower requirements. The estimated annual plant O&M costs derived from the data submitted by Vendor No. 3, and the MITRE assumptions taken from Table 4, are reasonably typical and are presented in Table 10.

The O&M costs shown in Table 10 do not include FGD waste disposal. Due to the densely urbanized locations of the power stations involved, there are no landfill sites available for

TABLE 10

ANALYSIS OF VENDOR QUOTATIONS: TYPICAL VALUES OF ANNUAL PLANT OPERATING AND MAINTENANCE COSTS FOR A NORTHEASTERN UTILITY - THOUSANDS OF 1982 DOLLARS (65 PERCENT OPERATING FACTOR)

Cost Item	Unit Cost	Plant A Units No. 1 & 2	Plant B
Sorbent (Lime of 90% purity)	$61/ton	2,370	2,870
Operating Labor and Supervision	$38,500/yr. (Fully burdened)	230	230
Process Water	$0.03/1,000 lb.	10	10
Sewer Charges (credit for use of dilution and spray down water)	$0.02/1,000 lb.	(50)	(70)
Electricity	$0.10/Kw-hr.	1,860	2,480
Compressed Air	$0.03/1000 scf	890	1,060
Maintenance Material (including filter bags, atomizer wheels, etc.)	$500,000/yr. for 80% operating factor	400	400
Labor	$38,500/yr. (fully burdened)	270	270
Analytical	$10/hr.	90	90
	Total Annual Cost	6,070	7,340
	Mills/Kw-hr.	1.29	1.51

disposal of the dry FGD waste (a mixture of approximately 2/3 fly ash and 1/3 spent sorbent), and of the bottom ash which would be generated from the conversion of the northeastern utility from oil-firing to low-sulfur coal-firing and from the installation of spray dryer FGD emission controls. Any waste disposal operations would have to be conducted at considerable distances from the utility sites. As a point of reference, if the total waste quantities generated were to be loaded into the empty rail cars that carried the coal from its West Virginia mine source to the northeastern utility power stations, a unit-train of waste could be shipped back to the mine for disposal every 3 to 4 days. Total transportation cost for this operation (assuming the waste freight tariff is identical to the coal freight rate to the utility sites), would be approximately $15/ton, exclusive of any costs associated with protection of the FGD waste from moisture in the open coal hopper cars.* Additionally, costs for placement of the waste in unused mine shafts and tunnels are unknown. The range of $10 to $20 per ton for total waste disposal cost (including FGD waste and coal ahs), while not based on site-specific analysis, is considered representative of the range of costs likely to be encountered by the northeastern utility for land disposal costs.** Table 11

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*The FGD waste as discharged from this system is essentially moisture-free (approximately 1 percent moisture) and contains significant amounts of unreacted lime. Admixture of appreciable amounts of water with this material could cause cementitious reactions to occur, resulting in the material "setting up" in the rail car.

**The northeastern utility believes that a range of $15 to $20/ton would be the likely range of waste disposal cost.

TABLE 11

PROPOSED SPRAY DRYER FGD SYSTEM ADDITION TO A NORTHEASTERN UTILITY PLANNING CONVERSION FROM OIL TO COAL-FIRING SOLID WASTE GENERATION - AMOUNTS AND ESTIMATED COSTS OF DISPOSAL[a]

WASTE TYPE	Plant A	Plant B	Total Both Plants
FLY ASH (TPY)[b]	177,600	228,800	406,400
BOTTOM ASH (TPY)	33,400	40,400	73,800
SPENT SORBENT (TPY)	85,800	104,000	189,800
TOTAL	296,800	373,200	670,000
ESTIMATED ANNUAL DISPOSAL COST, $ Millions (Mills/Kw.Hr.)[c]			
Unit Disposal Cost			
$ 10/TON	3.0 (0.6)	3.7 (0.7)	6.7 (0.7)
$ 20/TON	6.0 (1.2)	7.4 (1.4)	13.4 (1.4)

[a] At 65% load factor

[b] Tons Per Year - Fly ash and spent sorbent leave spray dryer FGD system as a combined stream

[c] 11% of these costs are for disposal of coal bottom ash which is assumed to be co-disposed with the dry FGD waste

presents estimated total solid waste disposal costs based on unit costs of $10 to $20 per ton. A comparison of total waste disposal cost at a unit cost of $20/ton with the estimate of plant O&M costs (Table 10) indicates that the waste disposal cost alone could be comparable to the total plant O&M cost.

Table 12 presents estimated first year annual revenue requirements based on capital cost estimates submitted by Vendors No. 1, 2 and 3 for the generalized northeastern location. Table 13 presents first year annual revenue requirements based on capital cost estimates submitted by Vendors No. 3 and 4 for the highly urbanized location. Total first year annual revenue requirements range from a low value of $44.5 million (4.5 mills/Kw-hr.) to a high value of $58.5 million (6.0 mills/Kw-hr.) for the spray dryer FGD systems installed in the two northeastern utility power stations in the generalized case. The higher total annual revenue requirement for the urban location (Vendor No. 3) averages $6.8 million (7.0 mills/Kw-hr.) over the range of FGD waste disposal costs used in this sutdy.

In order to develop potential savings over the cost of oil-firing, the total first year annual revenue requirements, reflecting installation of spray dryer FGD systems at the two northeatern utility power stations in an urban location, have been combined with annual coal costs (including the estimated coal waste disposal cost which would be associated with coal-firing whether or not FGD

TABLE 12

ANALYSIS OF FOUR VENDOR QUOTATIONS: ESTIMATES OF FIRST YEAR ANNUAL REVENUE REQUIREMENTS FOR A SPRAY DRYER FGD SYSTEM AT A NORTHEASTERN UTILITY - MILLIONS OF 1982 DOLLARS (MILLS/KW-HR.)[a]

COSTS APPLICABLE TO THE GENERAL NORTHEASTERN REGION

	Vendor No. 1		Vendor No. 2		Vendor No. 3	
	Power Plant A	Power Plant B	Power Plant A	Power Plant B	Power Plant A	Power Plant B
Direct Costs						
Plant Direct Costs	7.8(1.7)	8.7(1.7)	5.6(1.2)	6.3(1.2)	6.1(1.2)	7.3(1.4)
Waste Disposal Costs[b]	• 3.0(0.6)- 6.0(1.2)	3.7(0.7)- 7.4(1.4)	3.0(0.6)- 6.0(1.2)	3.7(0.7)- 7.4(1.4)	3.0(0.6)- 6.0(1.2)	3.7(0.7)- 7.4(1.4)
Total Direct Costs	10.8(2.3)-13.8(2.9)	12.4(1.4)-16.1(3.1)	8.6(1.8)-11.6(2.4)	10.0(1.9)-13.7(2.6)	9.1(1.8)-12.1(2.4)	11.0(2.1)-14.7(2.9)
Indirect Costs						
Overheads						
Plant Administrative (60% of Labor and Maintenance Charges)	0.9(0.2)	0.8(0.2)	—[c]	—[c]	0.9(0.2)	0.9(0.2)
Levelized Capital Charges (14.7% of total capital investment)	11.5(2.5)	12.3(2.4)	12.5(2.7)	13.4(2.6)	14.2(3.0)	15.7(3.1)
Total Indirect Costs	12.4(2.7)	13.1(2.6)	12.5(2.7)	13.4(2.6)	15.1(3.2)	16.6(3.3)
Total First Year Annual Revenue Requirements	23.2(5.0)-26.2(5.6)	25.5(4.0)-29.2(5.7)	21.1(4.5)-24.1(5.1)	23.4(4.5)-27.1(5.2)	24.2(5.0)-27.2(5.6)	27.6(5.4)-31.3(6.1)
Total First Year Annual Revenue Requirements (both plants)	48.7(5.0)-55.4(5.6)		44.5(4.5)-51.2(5.2)		51.8(5.3)-58.5(6.0)	

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[a] Costs at a 65% operating factor.

[b] These costs include the cost of disposal of approximately 89% FGD waste and 11% coal bottom ash, with the lower value based on $10/ton and the upper value based on $20/ton unit disposal cost.

[c] Overhead costs included in labor and maintenance portion of direct costs.

TABLE 13

ANALYSIS OF FOUR VENDOR QUOTATIONS: ESTIMATES OF FIRST YEAR ANNUAL REVENUE REQUIREMENTS FOR A SPRAY DRYER FGD SYSTEM AT A NORTHEASTERN UTILITY - MILLIONS OF 1982 DOLLARS (MILLS/KW-HR.)[a]

COSTS APPLICABLE TO A HIGHLY URBANIZED NORTHEAST AREA

	Vendor No. 3		Vendor No. 4	
	Power Plant A	Power Plant B	Power Plant A	Power Plant B[d]
Direct Costs				
Plant Direct Costs	6.1(1.2)	7.3(1.4)	5.1(1.1)	
Waste Disposal Costs[b]	3.0(0.6)- 6.0(1.2)	3.7(0.7)- 7.4(1.4)	3.0(0.6)- 6.0(1.2)	
Total Direct Costs	9.1(1.8)-12.1(2.4)	11.0(2.1)-14.7(2.9)	8.1(1.7)-11.1(2.3)	
Indirect Costs				
Overheads				
Plant Administrative (60% of Labor and Maintenance Charges)	0.9(0.2)	0.9(0.2)	--[c]	
Levelized Capital Charges (1.47% of total capital investment)	21.0(4.5)	22.5(4.4)	17.9(3.8)	
Total Indirect Costs	21.9(4.7)	23.4(4.6)	17.9(3.8)	
Total First Year Annual Revenue Requirements	31.0(6.5)-34.0(7.1)	34.4(6.7)-38.1(7.4)	26.0(5.5)-29.0(6.1)	
Total First Year Annual Revenue Requirements (both plants)	65.4(6.7)-72.1(7.3)		----	

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[a] Costs at a 65% operating factor.
[b] These costs include the cost of disposal of approximately 89% FGD waste and 11% coal bottom ash, with the lower value based on $10/ton and the upper value based on $20/ton unit disposal cost.
[c] Overhead costs included in labor and maintenance portion of direct costs.
[d] Data not available.

systems are added to the utility boilers) and with utility-provided estimates of the cost of modifications required to handle coal at the two power stations. Table 14 presents the assumptions used in developing this cost comparison and Table 15 presents the results of comparing oil and coal related costs. The results of the comparison show that a potential savings (1982 dollars) of $187 million (19 mills/Kw-hr.) annually can be expected for the two power stations by conversion from oil to low-sulfur coal, with particulate and SO_2 emissions maintained at present levels. Savings of this magnitude result in an attractive FGD system cost payback period, i.e., one to two years.

PROBLEMS ASSOCIATED WITH RETROFITTING SPRAY DRYER FGD SYSTEMS AT THE NORTHEASTERN UTILITY SITES

A number of considerations other than the cost of a spray dryer FGD system will affect the potential conversion of the northeastern utility oil-fired power boilers to coal-firing (with accompanying particulate and SO_2 emission controls). These factors include:

- availability of sufficient vacant land for the installation of the spray dryer FGD system
- disposal of FGD and coal wastes

Availability of Vacant Land for the Spray Dryer FGD System

As is indicated in Tables 6 and 7, Vendors No. 1, 2 and 3 supplied estimated total square footage required for installation of the spray dryer FGD system at Plants A and B, while Vendor No. 4 estimated the square footage needed at Plant A only. The average

TABLE 14

COST ASSUMPTIONS USED IN DEVELOPING COST COMPARISON OF OIL VS COAL PROPOSED CONVERSION OF NORTHEASTERN UTILITY (TWO POWER STATIONS) TO COAL FIRING

Basis: 1982 Dollars

Item	Value Used in Cost Calculations
Plant Operating Factor	65%
No. 6 Fuel Oil	$37/barrel delivered
Low Sulfur Coal	$55.50/ton delivered
Coal Ash Disposal Cost	$20/ton
FGD Waste Disposal Cost	$20/ton
Estimated Cost of Modification in Existing Facilities Required for Coal Firing (1980 Dollars)	
Plant A[a]	$48 million
Plant B[b]	$18 million

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[a] Not including the cost of special boiler modifications required for coal firing.

[b] Not including relocation costs for equipment and facilities required to provide sufficient area for the spray dryer FGD system, and realignment of existing ductwork to provide flue gas access to spray dryers.

TABLE 15

PROPOSED CONVERSION OF A NORTHEASTERN UTILITY TO COAL-FIRING WITH SPRAY DRYER FGD SYSTEM ADD-ON: COST COMPARISON OF OIL VS COAL FUELS (1982 DOLLARS)

Cost Item	Fuel-Related Costs in MM Dollars (Mills/Kw-hr.)		
	Plant A	Plant B	A and B Combined
Estimated Annual Oil Cost	229(49)	278(54)	507(52)
Estimated Annual Coal Cost (with Spray Dryer FGD system installed):			
Coal Cost and Associated Coal Waste Disposal Cost at $20/ton	112(24)	134(26)	246(25)
Modifications Required to Existing Plant in Order to Handle Coal	7(1.5)	2(0.4)	9(1)
Environmental Control Cost Include FGD Waste Disposal Cost at $20/ton (First Year Annualized Costs)[a]	31(7)	34(7)	65(7)
Total	150(32)	170(33)	320(33)
Estimated Annual Savings through Coal Use	79(17)	108(21)	187(19)

[a] Adjusted Vendor No. 3 for the urban situation.

area requirement for Plant A is approximately 115,000 square feet. There are approximately 300,000 square feet at Plant A available for installation of a spray dryer FGD system, with area available for some coal storage and some interim ash and FGD waste storage. Coal would be brought to the site by barge, and coal bottom ash and FGD waste would be removed by barge. Lime sorbent could be brought in by rail (approximately one and a half 100-ton hopper cars of lime would be required for each stream day).

The area requirements for a Plant B spray dryer FGD facility average approximately 109,000 square feet. Available vacant land at Plant B is quite limited with approximately 75,000 square feet of usable unencumbered area potentially available (currently an employee parking lot). Based on an on-site inspection, it is apparent that some equipment and facilities would have to be removed adjacent to the parking lot in order to provide the minimum block of continuous space needed for the spray dryer FGD installation. There would be no area available for coal or waste storage. However, barge delivery of coal and lime and barge removal of FGD waste and coal ash could be accomplished through suitable scheduling.* An analysis of the problems and costs involved in moving sufficient equipment, relocation of tankage, miscellaneous buildings, power lines, substations, etc., in order to provide adequate area for the spray dryer FGD installation, is outside the scope of this paper and is not addressed herein, although one should be prepared. An analysis should also be prepared of the environmental impact of the noise and fugitive dust created by the conversion of Plant B to coal-firing, since this plant is located in a densely populated urban area.

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*Plant B would be particularly vulnerable to coal mining and transportation strikes, since there is no raw material or waste holding capacity on-site.

Disposal of FGD Waste

A total of almost 3,000 tons per day of dry FGD waste and coal bottom ash would be generated by the northeastern utility Plants A and B. With no land available for waste storgae at Plant A and only a limited amount of interim storage area available at Plant B (perhaps one month storage), the need for landfill acreage for waste disposal would be one of the most pressing issues facing the utility in the proposed conversion to coal-firing. The urban locations of the two power stations precludes availability of sufficient landfill area in the immediate vicinity of the plants. Based on TVA data,* approximately 600 acres of landfill area will be required over the remaining life of the two plants. Land area of this size may not be available within 50 to 100 miles from the power stations. Rail transportation of waste to the coal mine source for disposal was discussed earlier. For FGD waste disposal in relatively nearby locations (within 50 miles) barging followed by transshipment to land transportation (rail, truck, or closed conveyor) to a number of relatively costly small land disposal sites* would be an alternate possibility.

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*EPA-600/7-80-060, pp. 17, 21, 40.

CONCLUSIONS.

An analysis has been performed to determine the costs and site-specific problems related to the retrofit of spray dryer FGD systems on a number of boilers located at two power stations of a northeastern utility. These retrofit would be in conjunction with conversion of these boilers from oil-firing to low-sulfur coal-firng. Conclusions drawn from this analysis are presented below.

- Based on budget estimates supplied by three vendors for a battery-limits spray dryer FGD facility, total installed costs for the two northeastern utility power stations involved range from $153.9 million ($89/Kw) to 204.0 million ($118/Kw) all in 1980 dollars. These are generalized costs for the northeastern region and are in the range of actual costs of spray dryer FGD systems presently being installed at three western utilities.
- Total installed costs in 1980 dollars for the two northeastern utility power stations located in a heavily urbanized area (typified by the northern New Jersey - southern New York area), are estimated to be $295.8 million ($171/Kw). This cost is based on budget estimates supplied by one vendor. Two other vendor estimates of a spray dryer FGD system for one of the power stations involved (based on an urban location) range from $121.7 million ($147/Kw) to $142.8 million ($173/Kw).
- Unit costs of spray dryer FGD waste disposal for the northeastern utility are expected to be in the $10 to $20 per ton range. Total solid waste generation at both plants is estimated to be 670,000 tons per year (at a 65 percent load factor).

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*TVA has used $5,000/acre for a landfill disposal site for a midwest location. Land values within 50 miles of the northeastern utility could be 10 times this value.

Total annual costs of disposal (including FGD waste, flyash and bottom ash) are estimated to range from \$6.7 million (1.3 mills/Kw-hr.) to \$13.4 million (2.6 mills/Kw-hr.). These costs represent 5 to 10 percent of the total annual revenue requirements attributable to the spray dryer FGD system operation.

- Total projected first year annual revenue requirements for operation of the spray dryer FGD system at the two northeastern utility power stations (including O&M costs, plant overhead, levelized capital charges, and the assumed range of waste disposal costs), range from \$44.5 million (4.5 mills/Kw-hr.) to \$58.5 million (6.0 mills/Kw-hr.) for the regional location case, and 65.4 million (6.7 mills/Kw-hr.) to \$72.1 million (7.3 mills/Kw-hr.) for the urban location case. All of these costs are based on a 65 percent load factor and in 1982 dollars.

- Total savings for the two northeastern utility power stations accrued through conversion from oil to low-sulfur coal-firing, with particulate and SO_2 controls achieved through the use of spray dryer FGD, are estimated to be \$187 million annually (19 mills/Kw-hr.)

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CURRENT STATUS OF DRY FLUE GAS DESULFURIZATION SYSTEMS

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Abstract

Radian Corporation is currently conducting a survey of the commercial and developmental status of dry FGD systems in the United States for the U. S. EPA. This paper will discuss the current commercial status of these systems, the focus of current research and development activities, the potential advantages of dry scrubbing over conventional wet scrubbing, and possible technical and economic limitations of dry FGD.

For the purpose of this study, dry FGD is defined as any flue gas desulfurization process producing a dry product for disposal. Dry FGD systems are grouped according to system type: (1) spray dryer based systems with ESP or fabric filter collectors, and (2) dry injection systems, primarily with baghouse collectors, and (3) other systems, including those where alkaline material is added directly to the fuel prior to combustion.

Of the three system types, only spray dryer systems have been commercially applied. Ten utility (low sulfur coal) systems had been sold as of May 1980. Two industrial spray dryer based systems have been sold. Higher sulfur coal (2 to 3 percent) is fired at the industrial sites. The dry injection/baghouse collection systems have been the subject of numerous past and ongoing bench and pilot scale studies, but no commercial systems have been sold to date. Technologies involving combustion of a coal/alkaline fuel mixture are still in the early stages of development. Two processes are currently under study: combustion of coal/limestone pellets and firing of a pulverized coal/limestone fuel mixture in a low-NO_x burner.

CURRENT STATUS OF DRY FLUE GAS DESULFURIZATION SYSTEMS

INTRODUCTION

Radian Corporation is currently conducting a survey of the commercial and developmental status of dry flue gas desulfurzation (FGD) systems in the United States. This project is being funded through the Environmental Protection Agency's (EPA) Industrial Environmental Research Laboratories. The paper presented today will discuss the current commercial status of dry SO_2 control systems and the focus of current research and development (R&D) activities. Also discussed are the possible advantages of dry systems vs. conventional wet lime/limestone systesm. Finally, the possible technical and economic limitations of dry systems are briefly addressed.

DEFINITION OF "DRY FGD SYSTEMS" CONSIDERED

For the purpose of this study, dry FGD is defined as any pollution control system where an alkaline material is contacted with SO_2-laden flue gas and a dry waste product results. This definition excludes fluidized bed combustion and several dry adsorption or "acceptance" processes, such as the Shell/UOP copper oxide process or the Bergbau-Forshung adsorptive char process.* The status of these programs is adequately discussed in other EPA reports.

Dry FGD systems can be grouped according to system type: (1) spray dryer based systems with ESP or fabric filter collectors, (2) dry injection systems, primarily with baghouse collectors, and (3) other systems, primarily those where alkaline material is added directly to the fuel prior to combustion such as a coal/limestone combustion system.

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*Rockwell's regenerable Aqueous Carbonate Process (ACP) was excluded as no solid waste product results. Rockwell has, however, adapted the open-loop portion of this process for a "throwaway" system.

In the spray drying process, flue gas is contacted with a slurry or solution such that the flue gas is adiabatically humidified and the slurry or solution is evaporated to apparent dryness. For FGD applications the sorbent is often a calcium-based slurry or a sodium solution which reacts with flue gas SO_2 during and following the drying process. The spray dryer can use rotary, two-fluid or nozzle atomization, and the vessel can be anything from the back-mix reactor typically used in conventional spray dryer applications to a large horizontal duct. The dried product salts and fly ash are collected in a downstream fabric filter or ESP.

Dry injection is defined as the process of introducing a dry sorbent into a flue gas stream. This can take the form of pneumatically injecting sorbent into a flue gas duct upstream of the particulate collection device, precoating or continuously feeding sorbent onto a fabric filter surface, or any similar form of mechanically introducing a dry alkaline sorbent to a flue gas stream.

Coal/limestone combustion is defined as the process of burning a mixture of coal and limestone whereby the SO_2 released from the coal reacts with the limestone to form solid calcium salts that are collected with the ash. Two specific combustion processes are currently being developed: one involves burning coal/limestone pellets in a stoker fired boiler, and the other involves burning a pulverized coal/limestone mixture in a low NO_x burner.

CURRENT STATUS OF DRY FGD TECHNOLOGY

Of the three system types discussed above, only spray drying has been commercially applied. There are currently 13 firms offering a commercial spray drying system. Ten utility and four industrial systems had been sold as of July 1980 (Table 1). Only two of these systems, at Celanese Fibers and Strathmore Paper, are operational. The first full

TABLE 1. KEY FEATURES OF COMMERCIAL SPRAY DRYING SYSTEMS SOLD TO DATE

System Purchaser/Vendor Utility	Location/Size	System Description	Coal	SO_2 Removal Guarantee	Reported Capital Cost	Reported Operating Cost	Status
Otter Tail Power Co./ Rockwell-Wheelabrator Frye.	Coyote Station, Beulah, ND/ Unit 1, 410 MW(1,890,000 acfm).	Four parallel spray towers, with 3 centrifugal atomizers each, followed by fabric filter with dacron bags. Will initially use commercial soda ash. Sorbent utilization guarantee of 80%.	North Dakota lignite; 0.78%S average; 7050 Btu/lb; 7% Ash.	70% for all fuels.	$32,000,000 ($78/kw)[a].	$ 6,580,000 ($2.5 mils/kwhr)[a]. Does not include waste disposal.	Start-up scheduled for mid-1981.
Basin Electric Power Coop/ Joy-Niro.	Antelope Valley, Beulah, ND/ Unit 1, 430 MW (2,200,000 acfm).	Five parallel spray dryers (one spare), single rotary atomizer per dryer, followed by fabric filter with teflon-coated fiberglass bags. Lime sorbent with partial recycle of solids. Ball mill slaker.	North Dakota lignite; 0.68% S average; 1.22% S maximum.	62% for average S coal; 78% for maximum S coal.	$49,665,100 ($113/kw)[b].	$2,270,834/yr ($0.8 mils/kwhr)[b]. Lime cost = $1,102,500 ($60/ton). Does not include waste disposal.	Start-up scheduled for April 1982.
Basin Electric Power Coop/ Babcock & Wilcox.	Laramie River, Wheatland, Wyoming/Unit 3, 500 MW (2,810,000 acfm).	Four parallel reactors (one spare) with 12 fluid nozzles each. Each reactor followed by an ESP. Lime sorbent, no solids recycle.	Wyoming subbituminous, 0.54% S average; 0.81% S maximum; 8140 Btu/lb; 8% Ash.	82% for average S coal; 90% for maximum S coal.	$49,807,000 ($83/kw)[b].	$2,571,000/yr ($0.7 mils/kwhr)[b]. Lime cost = $1,396,570 ($60/ton). Does not waste disposal costs.	Start-up scheduled for Spring 1982.
Northern States Power/ Joy-Niro. (2)	Riverside Station / Units 6 and 7, 110 MW (680,000 acfm); Retrofit.	One spray dryer with rotary atomizer. Will initially be demonstrated at 300,000 acfm with ESP. Full flow with fabric filter. Ball mill and attrition slaker for lime sorbent.	1% S Montana coal; 3.0 to 3.5% S Illinois coal; 4+% S coke.	Varying between 70 and 90% during demonstration tests.	Not available.	Not available.	Testing with ESP scheduled to start Fall 1980. Fabric filter on-line in early 1981.
Tucson Electric/ Joy-Niro.	Springerville Station/ Units 1 and 2; 350 MW each.	Spray dryer/fabric filter design. Lime sorbent. Rotary atomization.	New Mexico Coal; 0.69% S.	61%.	Not available.	Not available.	Unit 1 scheduled to start-up in late 1984; Unit 2 in 1986.
United Power Association/ Research-Cottrell.	Stanton Station, Stanton, Minnesota/65 MW.	Spray dryer/fabric filter rotary atomizers, possibly multiple atomizers per dryer. Lime sorbent.	Low and intermediate sulfur subbituminous Montana coal.	Not available.	Not available.	Not available.	Start-up scheduled for 1981.

*Multiply by 4.7 x 10^{-4} to convert acfm to m^3/s.

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TABLE 1. (Continued.) KEY FEATURES OF COMMERCIAL SPRAY DRYING SYSTEMS SOLD TO DATE

System Purchaser/Vendor Utility	Location/Size	System Description	Coal	SO_2 Removal Guarantee	Reported Capital Cost	Reported Operating Cost	Status
Platte River Power Authority/ Joy-Niro.	Rawhide Station/Unit 1 250 MW.	Spray dryer/fabric filter design. Rotary atomizers. Lime sorbent.	Western subbituminous coal; 1.3% S.	80%.	Not available.	Not available.	Start-up scheduled for 1983.
Colorado-Ute Association/ Babcock & Wilcox.	Craig Station/Unit 3 430 MW.	Horizontal spray dryers with nozzle atomizers, followed by fabric filter. Solids recycle. Ball mill slaker for lime sorbent.	0.70% S, 8950 Btu/lb, 14% Ash design coal; 0.40% S, 10250 Btu/lb, 8% Ash performance coal.	87% for design coal.	$45,000,000^c.	Not available.	Initial operation in November 1982. Commercial operation in April 1983.
Sunflower Electric Coop/ Joy-Niro.	Holcombe Station/Unit 1 310 MW.	Spray dryer/fabric filter. Rotary atomization. Lime sorbent.	Western subbitumnous coal.	80%.	Not available.	Not available.	Start-up scheduled for 1983.
Industrial							
Celanese Fibers Co./ Rockwell-Wheelabrator Frye.	Amcelle plant, Cumberland, Maryland/63,000 acfm (110,000 lb steam/hr).	Spray dryer with single rotary atomizer followed by fabric filter with felt/ figerglass bags. Paste slaker for lime sorbent. No solids recycle.	1.5% S and 2 to 2.5% S eastern coals.	70% for 1.5% S coal, 85% for 2.0% S coal.	$ 1,250,000^d.	Not available.	Operational. Passed Maryland state compliance tests in February 1980. Have achieved guaranteed removal.
Strathmore Paper Co./ Mikropul Inc.	Woronco, Massachusetts/ 40,000 acfm (85,000 lb steam/hr).	Spray dryer with four two-fluid nozzles, followed by fabric filter with specially finished acrylic bags.	2.3 to 3% S eastern coal.	75%.	$ 1,400,000^d,e.	$600/day^c;	Operational. Now achieving removal guarantee.
University of Minnesota/ Kennecott Development Co. (Environmental Products Division).	Univ of Minnesota/2 Units at 120,000 acfm each.	Spray dryer with single rotary atomizer followed by fabric filter with fiberglass bags. Lime sorbent.	Subbitumnous coal; 0.6 to 0.7% S.	70%.	$ 3,300,000^d,f.	Not available.	Commercial operation in Fall 1981.
Calgon/Joy-Niro.	37,000 acfm.	Spray dryer/fabric filter. Rotary atomizer. Soda ash sorbent. Removing SO_2 and HCl from 1700°F gases. Solids recycle.	1 to 2% S coal, 1100 ppm SO_2 + HCl.	75% SO_2; 90 % HCl	$ 1,600,000^d,g.	Not available.	Under construction.

*Multiply by 4.7×10^{-4} to convert acfm to m^3/s.

FOOTNOTES TO TABLE 1.

[a] Capital Cost for complete turnkey installation from air preheater outlet to stock connection, excluding I.D. fans (1977$). Source: Johnson, O.B., et. al., "Coyote Station, First Commercial Dry FGD Systems". (Presented at 41st Annual American Power Conference. Chicago, Illinois. April 23-25, 1979.)

[b] Evaluation based on 35-year plant life, capacity factor of 75% (1981$). Source: Janssen, K.E. and R.L. Eriksen. "Basin Electric's Involvement with Dry Flue Gas Desulfurization". (Presented at the EPA Symposium on Flue Gas Desulfurization. Las Vegas, Nevada. March 5-8, 1979.)

[c] Kelly, M.E. and S.A. Shareef. Meeting notes at Babcock & Wilcox. Barberton, Ohio. June 1980.

[d] Stern, J.L. "Dry Scrubbing for Industrial Flue Gas Desulfurization: State-of-the Art, 1980". (Presented at the 89th National Meeting of AICHE. Portland, Oregon. August 17-20, 1980.)

[e] From "ground-up". (1979$)

[f] "Straight through system". (1980$)

[g] (1980$).

scale new utility system is scheduled to start up in mid-1981 at Otter Tail Power Company's Coyote Station.

The utility systems are primarily for low sulfur western coal design requiring 70 to 85 percent SO_2 removal. The operating industrial installations treat gas from medium sulfur fuels (1.5 to 2.5 percent S). With the exception of one system, all the designs call for a lime sorbent: lime is less expensive than sodium alkalis and calcium based product salts may present less of a waste disposal problem than sodium-based salts. All but one of the systems offer a fabric filter for particulate collection, (an ESP is used in the utility system sold by B&W for Laramie River #3). Some vendors claim an additional 10 to 20 percent SO_2 removal across the fabric filter. However, vendors that offer ESPs claim that the gas can be cooled closer to saturation if an ESP is used instead of a fabric filter, resulting in additional SO_2 removal across the spray dryer.

The Northern States Power system is a retrofit design that will initially be operated as a demonstration unit by NSP and Joy/Niro. In addition to the NSP demonstration, there are several other large-scale demonstration units operating or being constructed (Table 2). Most of the demonstration test work involves investigation of various parameters such as sorbent type, waste solids recycle, inlet SO_2, sorbent stoichiometry, spray dryer outlet temperature, and waste solids properties. Some vendors are also privately conducting small-scale test work [0.47 to 2.6 M^3/s (1000 to 5000 acfm)] aimed at investigating the SO_2/sorbent reaction mechanism, effect of flue gas distribution in spray dryer, and the effect of fly ash alkalinity on SO_2 removal to help them better respond to bid requests. Other research efforts are focused on developing the spray drying process for high sulfur coal applications.

The dry injection/baghouse systems have been the subject of numerous past and on-going bench and pilot scale studies. Table 3 lists the

TABLE 2. MAJOR SPRAY DRYING DEMONSTRATION ACTIVITIES

Vendor and Utility	Babcock & Wilcox Pacific Power & Light	Buell-Envirotech/EPA Colorado Springs*	Combustion Engineering Northern States Power	Combustion Engineering Alabama Power	Ecolaire Systems Nebraska Power	Cottrell Environmental/EPA* Public Service of Colorado
Location	Jim Bridger Station	Martin Drake Station	Sherburne County	Gadsden Station	Gerald Gentleman Station	Comanche Station
Size**	120,000 acfm	8500 acfm	20,000 acfm	20,000 acfm	10,000 acfm	10,000 acfm
Status	Testing should be completed by Fall 1980.	Testing near completion.	Testing completed.	Tests to begin in the near future.	On-going tests.	Tests in progress.
SO_2 levels	-	-	1% S coal	up to 2000 ppm	300 to 1000 ppm	Low to medium sulfur Western coals.
Particulate collection	ESP and fabric filter.	Fabric filter.	ESP and fabric filter.	Fabric filter.	Fabric filter.	Fabric filter.
Sorbents	Lime, sodium-based sorbents.	Lime, limestone sodium-based sorbents adipic acid addition.	Lime.	Primarily lime.	Several types of lime and sodium carbonate.	Lime.
Atomization	V-jet nozzles.	Centrifugal atomizers.	Sonic atomization.	Sonic atomization.	Rotary atomizer.	Rotary atomizer.
Comments	-	Waste disposal studies also being conducted.	70% removal at stochiometric ratios of about 1 on once-through system.	Waste disposal studies to be carried out.	Future work may include tests with higher inlet SO_2 and nozzle atomization.	Expected to help define technical applicability limits.

* Results to be presented at this symposium.

** Multiply by 4.7×10^{-4} to convert to m^3/s.

TABLE 3. CURRENT DRY INJECTION PROGRAMS

Vendor/Agency	Location	Size	Comments
Buell Envirotech	Colorado Springs Martin Drake Station*	3000 acfm	Testing completed, May 1980. (EPA funded).
DOE/Grand Forks Energy Technology Center	GFETC Labs	200 acfm	Testing complete. Report expected in Fall 1980. Additional test work planned.
DOE/Pittsburgh Energy Technology Center	PETC Labs	500 lb coal/hr furnance	Testing completed.
EPRI/KVB	Public Service Company of Colorado-Cameo Station	20 MW_e	Tests underway.

* To be reported during this symposium.

current dry injection programs. No commercial systems have been sold to date and few vendors even offer a commercial dry injection system. Fairly high temperatures (>600°F) are required to achieve significant SO_2 removal using lime or limestone. The most reactive sorbents are sodium based (nacholite or trona), resulting in waste disposal problems. Nahcolite has been demonstrated to be more reactive than trona but has not been available in commercial quantities.

Technologies involving combustion of a coal/limestone fuel mixture are still in the early stages of development. Two of the most promising systems are combustion of coal/limestone pellets and firing of a pulverized coal/limestone fuel mixture in a low-NO_x burner. EPA is sponsoring development of both these technologies.

A 30-day test program with the coal/limestone pellets was scheduled to start in September 1980. The tests will be conducted at General Motor's Indianapolis plant (60,000 lb steam/hr stoker boiler). This technology is still in the developmental stage and has only been tested on small boilers to date. An accelerated program to demonstrate this technology is currently underway, including economic and vendor capability studies.

The SO_2 removal effectiveness of the coal/limestone mixture fired in a low-NO_x burner is apparently related to the lower flame temperatures present in the low-NO_x burner. The relatively high flame temperatures in conventional burners may result in "glazing" of the reagent particles, leading to significantly lower reactivity.

DRY FGD VS. CONVENTIONAL WET (LIME/LIMESTONE) SCRUBBING

Technology comparisons between dry and wet scrubbing systems can be drawn in several major areas: waste disposal, reagent requirements, operation and maintenance, energy requirements, and economics. This comparison will focus on general aspects of dry FGD systems as compared to conventional lime/limestone wet scrubbing systems.

With regard to waste disposal, dry FGD systems have an inherent advantage over wet lime/limestone systems in that they produce a dry, solid waste product that can be handled by conventional fly ash handling systems, eliminating requirements for a sludge handling system. However, the waste solids from sodium-based dry FGD systems are quite water soluble and can lead to leachability and waste stability problems. Waste solids from lime spray drying systems and coal/limestone fuel systems should have similar environmental impacts as waste from wet lime/limestone systems, for which waste disposal technology is better defined.

In general, dry FGD systems require a significantly higher stoichiometric ratio of sorbent to entering SO_2 to achieve the desired removal efficiency than do conventional limestone wet scrubbing systems. In addition, the reagents used in spray drying and dry injection systems (soda ash, lime, commercial and naturally occurring sodium carbonates and bicarbonates, such as nahcolite and trona) are significantly more expensive than limestone. Consequently, wet limestone scrubbing systems will have an advantage with regard to both reagent utilization and sorbent-related operating costs. It should be noted that, in comparing reported stoichiometric ratios for dry and wet systems, stoichiometry for the dry system is often reported as moles sorbent required per mole of SO_2 entering, while the conventional basis for wet systems is moles sorbent required per mole of SO_2 removed.

Several vendors claim that dry systems will have lower maintenance requirements than comparable wet systems. Dry systems require less equipment than wet systems because the thickeners, centrifuges, vacuum filters and mixers required to handle the wet sludge waste product from wet systems are eliminated. In addition, slurry pumping requirements are much lower for spray drying and are eliminated in dry injection and combustion of coal/alkali fuel systems. This is important because wet systems have reported high maintenance requirements associated with large slurry circulation equipment. Finally, the scaling potential in wet limestone systems requires extra effort to maintain proper scrubber operation and possibly makes dry systems somewhat more flexible as far as their ability to adjust process operations to respond to variations in inlet SO_2 concentrations and flue gas flow rates.

With regard to energy requirements, dry FGD systems appear to have a significant advantage over wet systems due to savings in reheat and pumping requirements. Spray dryer systems are usually designed to achieve required SO_2 removal while maintaining a 16°C to 36°C (30° to 50°F) approach to the adiabatic saturation temperature of the gas at the outlet of the spray dryer. Some systems are designed with warm gas (downstream of the air preheater) or hot gas (upstream of the air preheater) bypass. An energy penalty is associated with the use of hot gas bypass, and reduced overall SO_2 removal results from the use of warm or hot gas bypass.

Energy savings from reduced pumping requirements result from the fact that wet scrubbers may require liquid-to-gas ratios (L/G) pumping rates of up to 100 gallons per 1000 acfm whereas the L/G for spray drying systems ranges from 0.03 to 0.04 liters/m (0.2 to 0.3 gallon per 1000 ft^3).

One of the major driving forces for development of dry SO_2 removal systems is the opportunity for reduction in both capital and operating costs relative to conventional wet systems. Although costs are quite site specific, the three types of dry FGD technologies considered here offer several potential possibilities for cost savings. This is due to the reduction in equipment and operation and maintenance requirements relative to conventional wet lime/limestone systems, especially in utility applications. Basin Electric evaluated the costs of the spray drying systems purchased for the Antelope Valley and Laramie River Stations to be about 20 to 30 percent less over the 35-year life of the plant than comparable wet systems. However, it should also be noted that these economics are based on pilot scale data and should be better determined after the operation of commercial systems has begun. Dry systems may lose this advantage with respect to operating costs for high sulfur applications due to their higher sorbent requirements.

The minimal equipment and operating requirements for dry injection systems make the process economically attractive as far as capital costs are concerned, but high sorbent requirements and uncertainties in sorbent availability and cost are slowing further development of the technology on a commercial scale. Capital costs for both the pellet and low-NO_x burner coal/limestone fuel mixture systems should also be low since they will consist mainly of the equipment needed to produce the mixtures. However, since these systems have the potential for impacting the design and/or operation of the boiler, more information on the overall operability of these systems is needed before total operating costs can be estimated.

TECHNICAL AND ECONOMIC LIMITATIONS OF DRY FGD

Spray Drying

The application of spray dryer technology to higher sulfur coals may be subject to economic and technological limitations. Higher stoichiometries are required for higher sulfur applications. Consequently, the reagent cost differential between lime and limestone may alone make a spray dryer-based system uneconomical. The inlet flue gas temperature and the required overall SO_2 removal efficiency limit the amount of solution or slurry that can be sprayed into the gas. Thus for high sulfur loadings where high stoichiometries are required, it may be difficult to achieve high SO_2 removals. There is also an upper limit on the solution concentration or weight percent solids slurry that can be achieved.

Dry Injection

Major restraints to the development of dry injection technology have been uncertainty in sorbent (nacholite) availability, waste-disposal problems associated with sodium-based salts, and the relatively high flue gas temperatures required to achieve high SO_2 removals with calcium-based sorbents. Current research is focusing on improving the reactivity of more available sorbents such as trona and developing suitable solids disposal methods.

Combustion of Coal/Limestone Fuel Mixture

To date, only preliminary data exist for the SO_2 control effectiveness and operation of boilers firing either coal/limestone pellets or a pulverized coal/limestone fuel mixture. Further research on a larger scale for both systems is needed to determine the effects of firing a coal/limestone fuel mixture on boiler operation and maintenance. Effects of the increased particulate loading and the degree of SO_2 removal achievable also need to be investigated further.

SUMMARY

Dry flue gas desulfurization is a rapidly developing technology, due in part to the potential advantages it offers over conventional wet scrubbing techniques, especially for low sulfur coal applications. Spray drying is the only commercially applied technique but research and development work is continuing on both dry injection and combustion of fuel/alkali mixtures. Research efforts are focused on gaining a better understanding of the reaction mechanisms, characterizing waste solids properties, and developing techniques to increase process applicability.

Acknowledgements

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DRY SO_2 SCRUBBING PILOT TEST RESULTS

by

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ABSTRACT

Dry SO_2 scrubbing has emerged as an attractive technology for flue gas desulfurization to meet NSPS. Pilot testing by Research-Cottrell and others indicates that a spray dryer followed by a particulate collection device is a viable alternative to wet scrubbing for low sulfur coals.

Research-Cottrell and its spray dryer supplier, Komline-Sanderson, have conducted several investigations in Texas and Colorado employing spray dryer/fabric filter technology. This paper describes some of the results of the Research-Cottrell/Komline-Sanderson pilot test work performed for its commercial interests as well as for an EPA-funded program. Key process parameters that affect SO_2 removal, including stoichiometry, flue gas temperature, inlet SO_2 concentration and recycle, are discussed. Other related variables and the limits of dry SO_2 scrubbing are also considered.

DRY SO_2 SCRUBBING PILOT TEST RESULTS

This paper is an interim report of the results of dry SO_2 scrubbing test programs by Research-Cottrell and its exclusive spray dryer supplier, Komline-Sanderson, employing a spray dryer/fabric filter pilot system. The information presented is derived from field pilot plant programs conducted at the Comanche Station of Public Service of Colorado and at the Big Brown Station of Texas Utilities. Funding for the test program is from Research-Cottrell/Komline-Sanderson and the United States Environmental Protection Agency. The EPA-funded test program is currently in progress at Comanche Station, Pueblo, Colorado.

PILOT PLANT DESCRIPTION

The pilot test system (see Figures 1 and 2) is designed to treat 10,000 ACFM (nominal) of flue gas. The system consists of a spray dryer to remove SO_2 followed by a fabric filter to collect dry FGD solids and flyash as well as remove additional SO_2. An induced draft fan moves the flue gas through the system and a second "reverse air" fan cleans the fabric filter. Feed tanks and metering pumps are provided to supply reagent to the system. An SO_2 tank and delivery system provide additional SO_2 to the flue gas for tests at higher inlet SO_2 concentrations. Ductwork and dampers for flue gas transport and diversion are part of the system. In keeping with normal pilot plant practice, most control is accomplished manually in order to maintain maximum operating flexibility. The unit is thoroughly instrumented to provide the necessary basis for control as well as to provide complete detailed data to define the performance of the unit.

Spray Dryer

The pilot spray dryer is an 8'Ø x 35' high unit equipped with a variable speed rotary disc atomizer. Dirty flue gas containing SO_2 and flyash enters the top of the spray dryer where it is intimately

Figure 1 Research-Cottrell/Komline-Sanderson Dry Scrubbing Pilot Plant.

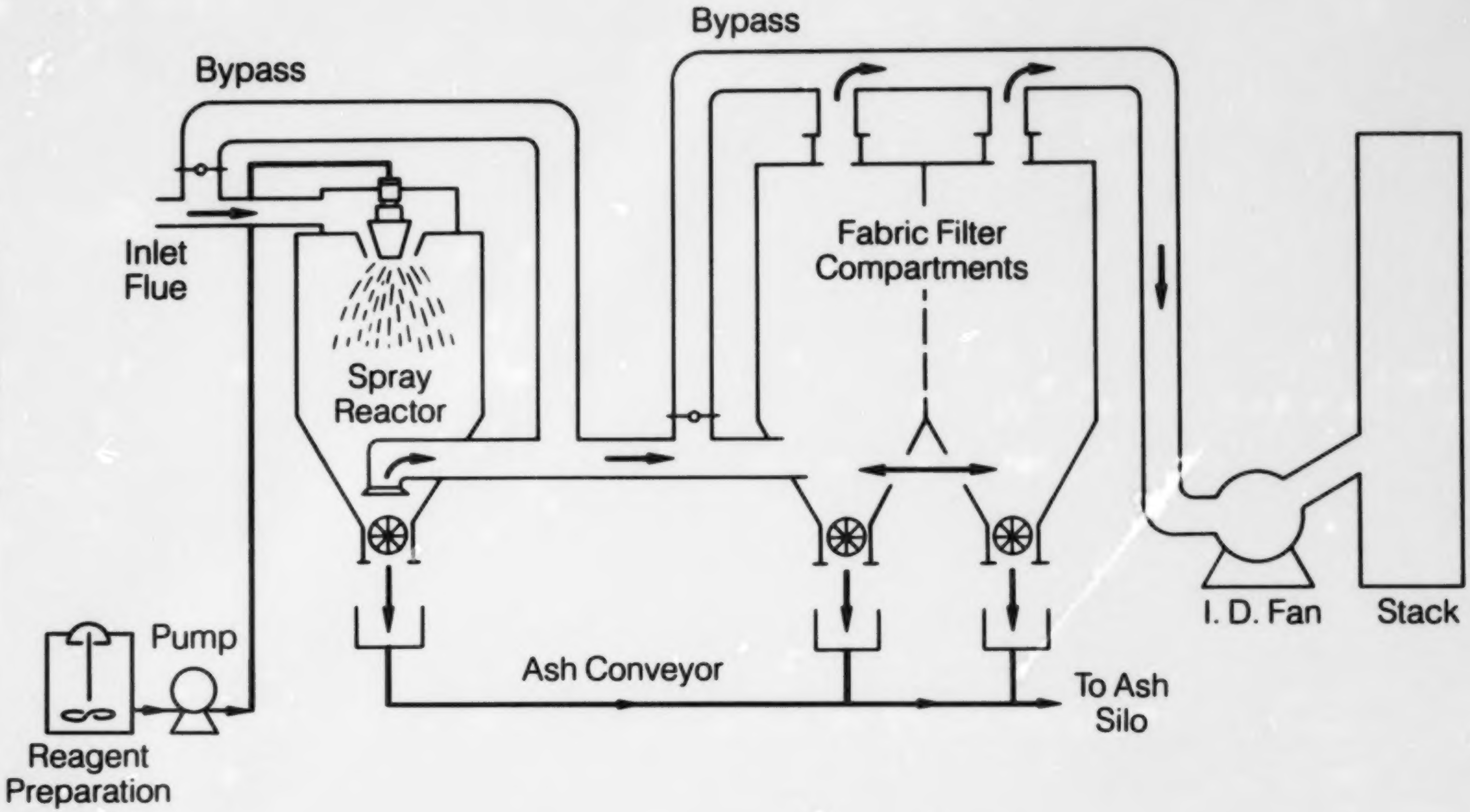
Bypass
Bypass
Inlet Flue
Spray Reactor
Fabric Filter Compartments
I. D. Fan
Stack
Pump
Reagent Preparation
Ash Conveyor
To Ash Silo

Figure 2 Research-Cottrell/Komline-Sanderson Dry Scrubbing Pilot Flow Diagram.

contacted with finely atomized lime slurry. The intimate contact and large interfacial area provided by the spray dryer result in very rapid SO_2 absorption by the lime slurry. Most of the SO_2 removal in the overall system occurs in the spray dryer. Before leaving the dryer, the solids approach complete dryness. In the pilot unit, coarse solids tend to settle in the conical bottom of the dryer and are discharged through a rotary valve to receiving drums. The scrubbed flue gas containing finer particles leaves the dryer through a side exit port and flows to the fabric filter. The construction of the spray dryer permits considerable flexibility in location of the gas outlet position and degree of solids removal as well as total contact time.

Alteration of the atomizing disc rotating speed, to achieve better atomization or lower power requirements, is achieved by replacing drive components. During testing, conditions are held at fixed values until the system operation responses, particularly sulfur dioxide removal efficiency and temperature drop, are accurately measured. Primary, quick response control of the spray dryer is obtained by varying the total lime feed rate and the total water feed rate. Temperature drop across the unit is controlled by total water flow to the unit. Water for temperature control is metered as a trim water flow to a mixing tee immediately upstream of the spray dryer atomizer. Increasing lime reagent flow increases the SO_2 removal efficiency at a given water flow, gas flow and sulfur dioxide content. System responses to changes in stoichiometry are observed on a duPont UV analyzer which samples both inlet and outlet streams from the spray dryer and fabric filter.

Fabric Filter

The fabric filter unit size is 10' x 15' x 55' and it contains two-sixteen bag compartments each designed to process about 5,000 cfm of flue gas. Each of the fabric filter bags is of commercial utility dimensions, 12" Ø x 30' high, and contains about 94 ft^2 of bag surface.

Flue gas from the spray dryer continuously enters the bottom of the fabric filter unit and leaves from the top. In the fabric filter, flyash and FGD solid particulate are removed from the flue gas and additional SO_2 removal takes place. From the fabric filter, the flue gas flows to the I.D. fan and then to the stack.

Although the majority of SO_2 removal takes place in the spray dryer, an important percentage of SO_2 removal occurs in the fabric filter. During an operating cycle, a thin layer of solid particles containing unreacted lime continually builds up on the surface of the filter bags. Flue gas at $140^{\circ}F$-$170^{\circ}F$ and with reduced SO_2 concentration flows through the bed of finely divided solids containing unreacted reagent. Gas-solids contact is good and some additional SO_2 removal occurs although the gas contact time in the bed is short.

To limit pressure drop across the filter bags from the accumulation of collected solids, a flow of air periodically is passed through the bags in the reverse direction for a short period of time (one-two minutes per hour). The "reverse air" flow causes most of the solids deposited to dislodge from the bag surface and drop into the collection hopper where they are discharged through rotary valves. During the brief bag cleaning period, flue gas is bypassed around the fabric filter. Valve operation is accomplished from the control room by either manual valve direction or by use of an automatic sequencing system.

Slurry Feed System

Pebble or granular lime is fed from the lime bin to the Vibra-Screw feeder. The feeder drops the quick lime into the pugmill slaker where it is mixed and reacted with water. The hydrated lime slurry is diluted, degritted and fed by gravity into the lime tank. The rate and concentration of lime slurry produced are controlled by adjusting the Vibra-Screw lime rate and the water rate to the slaker.

To maintain uniformity, lime slurry is recirculated around the lime tank by means of the lime pump. A bleed stream from this tank is metered and pumped directly to the spray dryer tank or to the recycle tank depending on the mode of operation.

When high SO_2 removal is desired, reagent use becomes excessive unless solids recycle is practiced. These solids are obtained from the baghouse and/or spray dryer and contain unused lime values. The solids are dumped into the hopper of the recycle tank where the fresh lime is added. Slurry from this tank is metered into the spray dryer atomizer. Alternative operation is to feed the recycle solids to a ball mill where they are mixed with water and ground to smaller particles sizes before entering the recycle tank.

PILOT TEST PROGRAMS

The pilot plant field test programs were executed at power plant sites in Colorado and Texas to determine the key process parameters that affect SO_2 removal in the spray dryer/fabric filter system. These empirical investigations were primarily to identify the dominant variables and to establish their importance rather than to determine why they are important or the underlying mechanisms involved.

The pilot tests have consisted of three different types of activities:

- o Process variable studies using lime reagent in once-through operation with no recycle. Key spray dryer and fabric filter process variables were studied to establish their relative importance with respect to SO_2 removal. Additional continuous tests were conducted to further define the effects of the operating variables and to test different atomizer discs. The pilot tests were conducted either in 12-16 hour shifts/day or in "around-the-clock" operation.
- o Process variable studies incorporating recycled solids. Continuous process variable tests utilizing solids re-

cycled from the spray dryer and fabric filter were carried out to remove SO_2 with improved lime reagent utilization.

- Longer term continuous, around-the-clock demonstration runs. The continuous demonstrations utilizing recycled solids were one-two weeks in duration and demonstrated smooth spray dryer/fabric filter system operation without significant mechanical problems.

Process conditions were varied during the demonstration runs to simulate variations that might be encountered during commercial operation and to permit acquisition of data at many different conditions. This information contributed significantly to the total data base generated in the pilot test programs.

The process variables considered relevant to SO_2 removal investigated during the pilot testing are presented in Table 1. These parameters were studied over the ranges of conditions shown in Table 2. Numerous additional operating conditions were observed and recorded during the pilot tests to assure controlled test operations.

Table 1 VARIABLES INVESTIGATED

SPRAY DRYER

- Lime/SO_2 Stoichiometry
- Inlet SO_2 Concentration
- Inlet Gas Temperature
- Outlet Gas Temperature
- Flue Gas Rate
- Atomizer Speed
- Lime Slaking Conditions
- Solids Recycle

FABRIC FILTER

- Air-to-Cloth Ratio
- Temperature
- Pressure Drop
- Stoichiometry
- Inlet SO_2 Concentration

Table 2 RANGE OF PILOT TEST VARIABLES

VARIABLE	MINIMUM	MAXIMUM
Inlet SO_2 Concentration, PPMV	400	2000
Stoichiometry, Moles CaO/Moles SO_2 in	0.4	5.2
SD Inlet Flue Gas Temperature, ^oF	195	340
SD Outlet Flue Gas Temperature, ^oF	125	210
Fabric Filter Gas Temperature, ^oF	117	200
Atomizer Disc Diameter, in.	7-3/8	8-1/2
Atomizer Disc Speed, RPM	10,000	14,000
Inlet Flue Gas Rate, ACFM	3,000	7,500
Fabric Filter Air-To Cloth Ratio, ft/min.	1.3	2.5
Lime Slurry Feed Conc., wt.%	3	25
Recycle Slurry Solids Conc., wt.%	10	53
Recycle Ratio, lbs recycle/lbs make up lime	0	6

Pilot test data were analyzed using a combination of visual inspection of graphical presentations and computer regression analysis. The independent variables that most strongly influence SO_2 removal in the spray dryer and in the fabric filter were arrived at by this means.

SPRAY DRYER PILOT RESULTS

External Variables

The dominant external variables in the spray dryer were determined by using a computer-aided, step-wise multiple regression analysis (MUL-CORRELATION Routine Of STATSYST*** Program). The variables in order of decreasing impact for SO_2 removal efficiency are:

- o Stoichiometric Ratio (Strong)
- o Temperature Approach to Adiabatic Saturation (Strong)
- o Temperature Drop Across Spray Dryer (Moderate)
- o Flue Gas Inlet SO_2 concentration (Moderate)

Flue gas rate had a negligible effect on SO_2 removal in the range investigated.

Stoichiometric Ratio (S.R.). As expected, lime stoichiometry significantly affects SO_2 removal in dry scrubbing. SO_2 removal efficiency increases strongly as S.R. is increased and tends to level off at higher S.R. values. From Figure 3, it appears that SO_2 removal increases nearly linearly as S.R. is increased at large values of temperature drop across the dryer but levels off much more quickly at lower temperatures differences. More efficient lime utilization might be expected at higher temperature differences since the temperature decrease across the spray dryer is essentially directly proportional to the quantity of water added with lime slurry feed. At otherwise identical conditions, increased water should enhance SO_2 removal. The data shown in Figure 3 were obtained at spray dryer flue gas temperatures far enough removed from adiabatic saturation that this temperature effect was minimized.

Stoichiometric ratio is defined as moles slaked lime/mole SO_2 fed to the spray dryer. In this paper, S.R. is reported on a pure slaked lime basis. That is, the total lime is corrected for lbs. $Ca(OH)_2$ obtained by titration with 1N HCL to a phenolphthalein end point/lb. solids in an analyzed lime slurry sample. Stoichiometric ratios reported on this active or available lime basis are 85-90 wt.% of S.R. values arrived at by employing the total weight of lime fed.

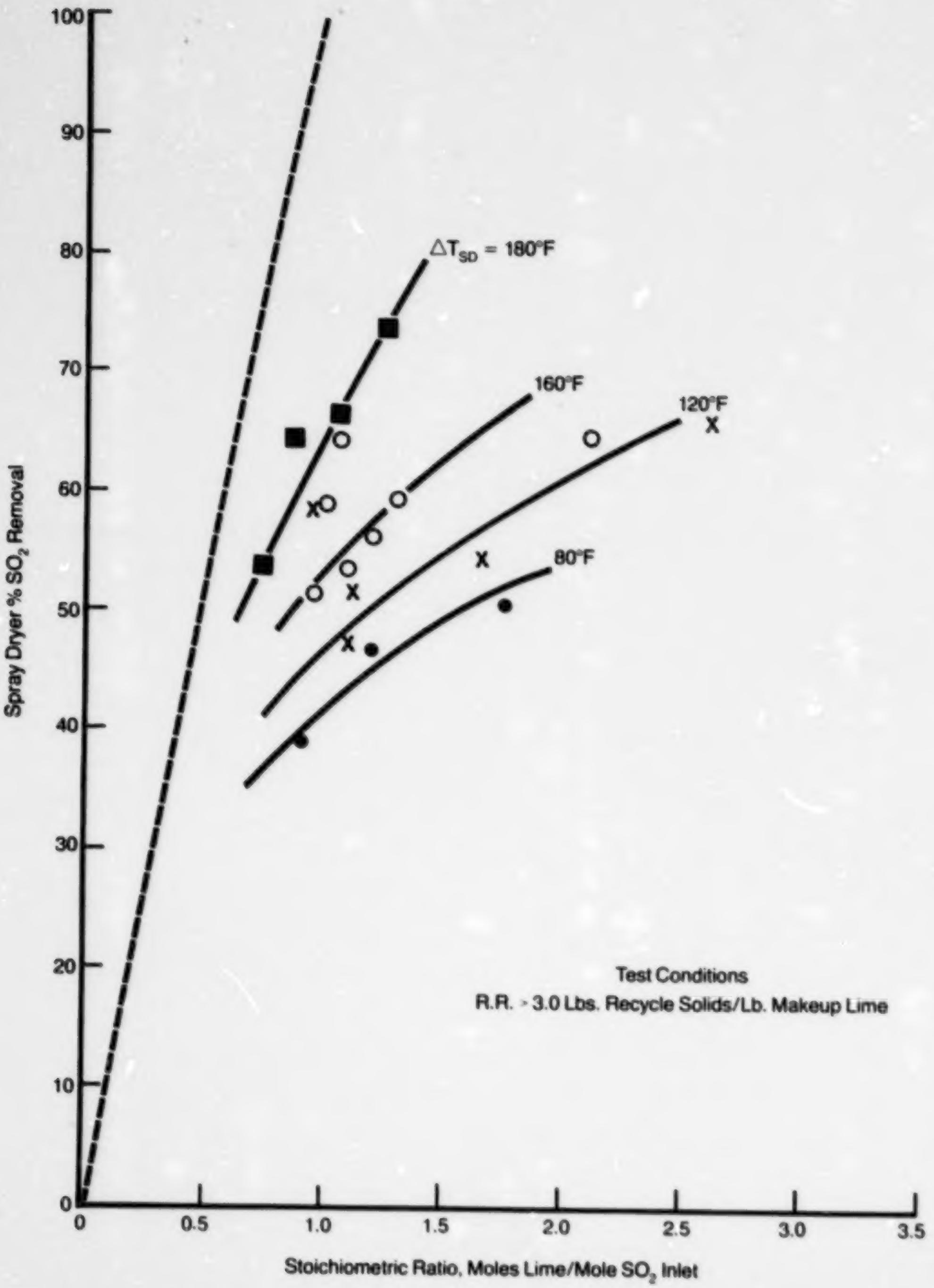

Figure 3 Effect of Stoichiometry on Spray Dryer SO_2 Removal.

Temperature Approach to Adiabatic Saturation (ΔT_{AS}). SO_2 removal increases with decreasing flue gas temperature in the spray dryer. The spray dryer bulk flue gas temperature was found nearly isothermal and virtually identical to the outlet temperature. SO_2 removal increases sharply as the flue gas temperature approaches the adiabatic saturation temperature, as shown in Figure 4. The greatly improved SO_2 removal is apparently a result of the increased moisture retained by the slurry as it dries. Solids collected from the spray dryer bottoms show significantly increased moisture content as the flue gas temperature approaches adiabatic saturation (Figure 5). The increases in SO_2 removal and in spray dryer solids moisture content closely parallel each other.

As the flue gas temperature is decreased, SO_2 removal increases but the solids moisture content increases to the point where difficulties arise in discharging solids from the spray dryer bottom. Thus, the optimum operating temperature strikes a balance between high SO_2 removal and trouble-free, continuous operation from the discharge and subsequent handling of high moisture content solids.

Temperature Drop Across Spray Dryer (ΔT_{SD}). The flue gas temperature drop across the spray dryer affects SO_2 removal in the spray dryer. Increased SO_2 removal is experienced as ΔT_{SD} is increased, according to Figure 3. Since flue gas temperature drop across the spray dryer is achieved by the evaporation of the water supplied, the temperature drop is directly related to the liquid/gas ratio (L/G) fed to the spray dryer. (L/G is also an important variable that affects SO_2 removal in wet FGD scrubbers).

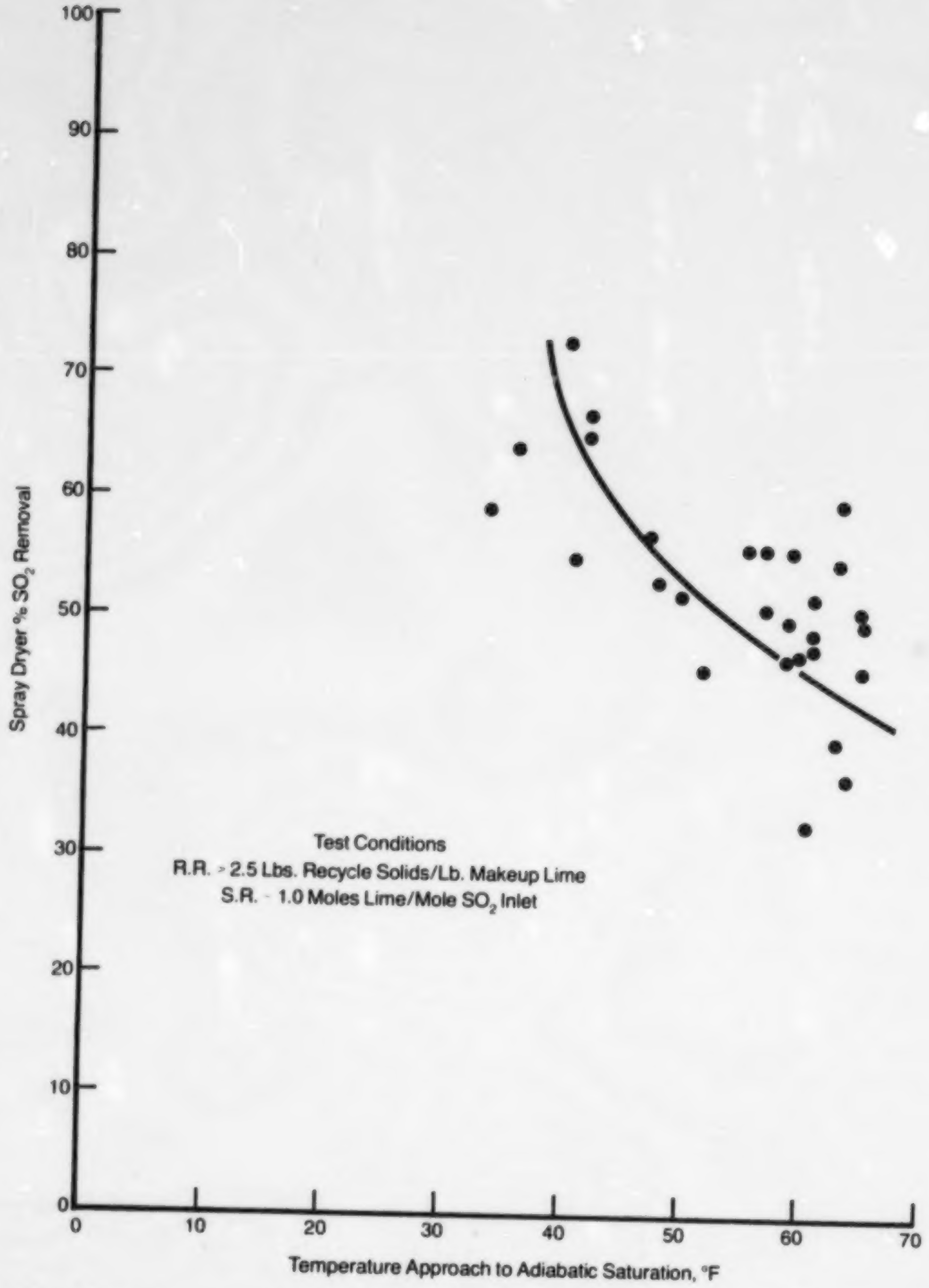

Figure 4 Effect of Flue Gas Temperature Approach to Adiabatic Saturation on Spray Dryer SO_2 Removal.

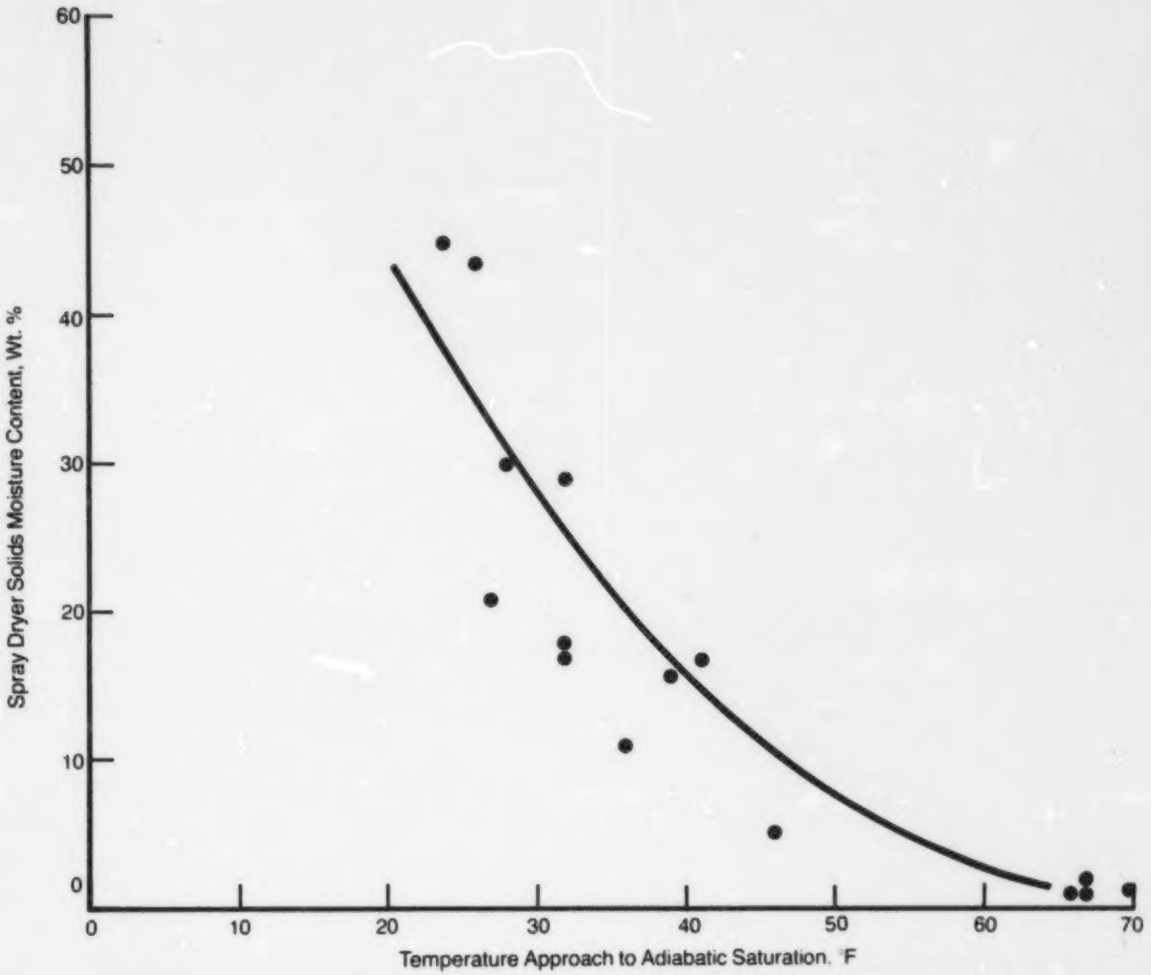

Figure 5 Effect of Flue Gas Temperature Approach to Adiabatic Saturation on Spray Dryer Solids Moisture Content.

Inlet SO_2 Concentration. SO_2 removal in the spray dryer decreases gradually with increasing flue gas inlet SO_2 concentration in the range studied from 400 to 2000 ppm SO_2. As Figure 6 indicates, SO_2 removal decreases with increasing ppm SO_2 at about the same rate at different levels of stoichiometry. The effect of SO_2 concentration on SO_2 removal is similar whether recycle or once-through lime operation is employed.

Flue Gas Rate. Flue gas rate was observed to have no significant effect on spray dryer SO_2 removal. Flue gas rate is inversely related to the gas phase residence time in the spray dryer. In the range of 5 to 10 seconds used in the pilot studies very little, if any, change in SO_2 removal was observed as gas phase residence time was studied by varying flue gas flow to the spray dryer.

Other Parameters

Atomizer Speed. Good SO_2 removal was observed at 10,000 RPM and at 14,000 RPM atomizer disc speeds using approximately 8" diameter discs in the pilot tests. Increasing atomizer speed produces a dryer solids product, especially at higher flue gas throughputs. In one test, wet product solids were produced in the spray dryer and fabric filter at the 10,000 RPM disc speed and 6,000-8,000 ACFM flue gas flow to the spray dryer. Except for the runs noted here, all of the data reported in this paper was obtained at an atomizer speed of 14,000 RPM.

The rotary disc atomizer consists of a titanium body with six silicon carbide ports inserted around the periphery. The bottom plate of the titanium disc is coated with aluminum oxide for protection against slurry abrasion.

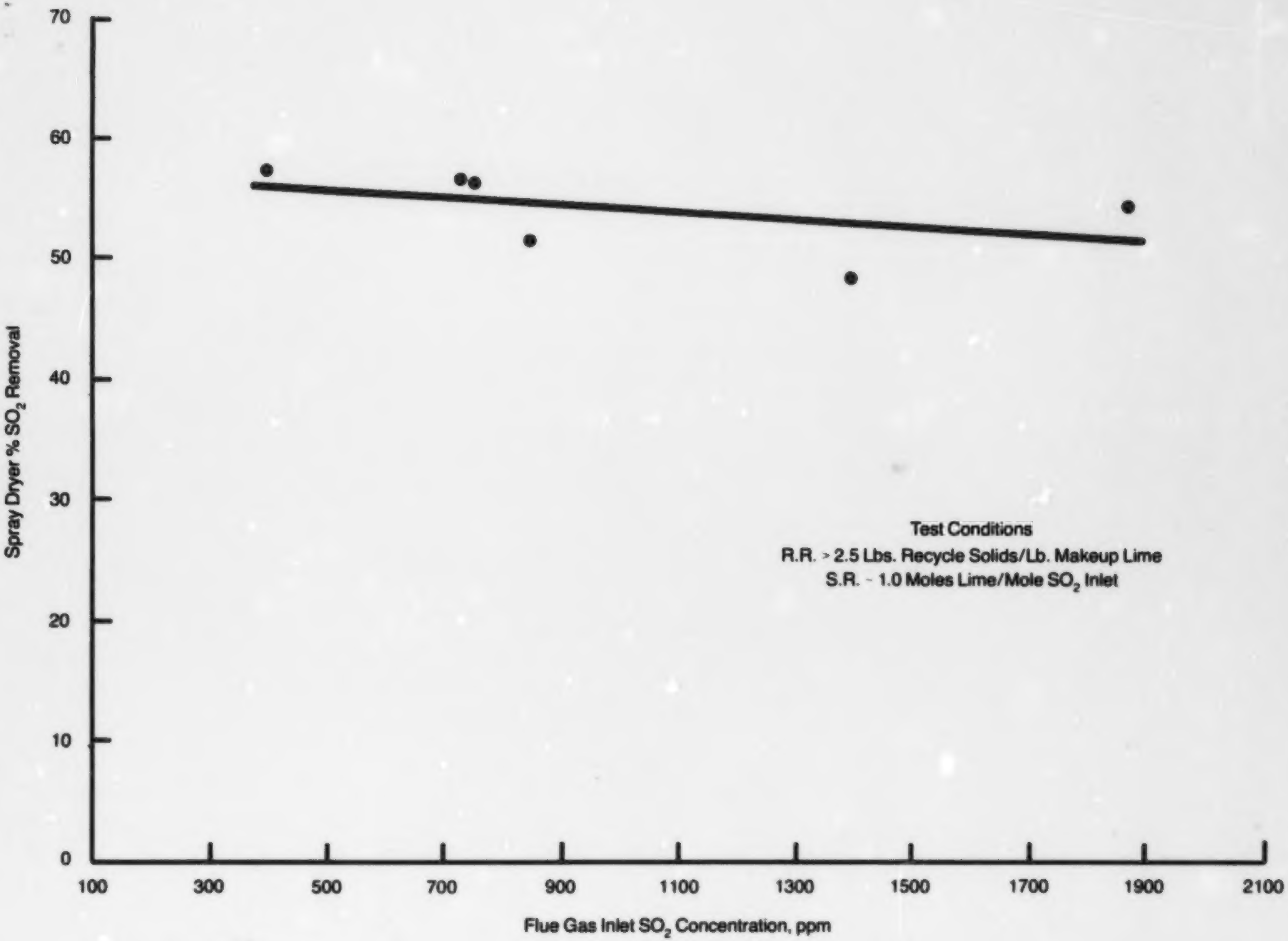

Figure 6 Effect of Inlet SO_2 Concentration on Spray Dryer SO_2 Removal.

Lime Slaking Conditions. Optimum conditions for maximum lime reactivity and minimum particle size were based upon consultations with lime slaker vendors. Water addition to the pugmill slaker was controlled to achieve a slaker temperature of 185°F during the majority of the tests. Operation at a slaker temperature of 170°F produced no noticeable change in SO_2 removal. Untreated process water was used as slaking water and also for the remainder of the makeup water required.

Solids Recycle. Recycling the mixture of FGD product and flyash solids to the spray dryer significantly enhances SO_2 removal efficiency and improves fresh lime utilization. Figure 7 shows that SO_2 removal in the spray dryer increases with increasing recycle ratio up to 2.5-3.0 lbs. recycle solids/lb. makeup lime where it levels off. At S.R. = 1.0, SO_2 removal increases from 53% with no recycle (Recycle Ratio = 0) to a maximum of 62%. Figure 7 also shows that SO_2 removal is not significantly different employing either spray dryer solids or fabric filter solids as the recycled material.

Improved SO_2 removal employing product solids recycle occurs both in the spray dryer and fabric filter. While SO_2 removal is greater using recyle, the general effects and trends of the key process parameters - stoichiometry, flue gas temperature, flue gas SO_2 concentration - are similar with either recycle or once-through lime operation.

The considerable improvement in SO_2 removal observed with recycle is greater than can be accounted for based on the additional lime content of the recycle solids. Electron micrographs suggest that the lime present is made more available for reaction because of the increased area for dispersion on the recycle solids. In limited testing, ball milling of the recycle solids before reintroduction into the spray dryer produced minimal improvement in SO_2 removal over unmilled solids.

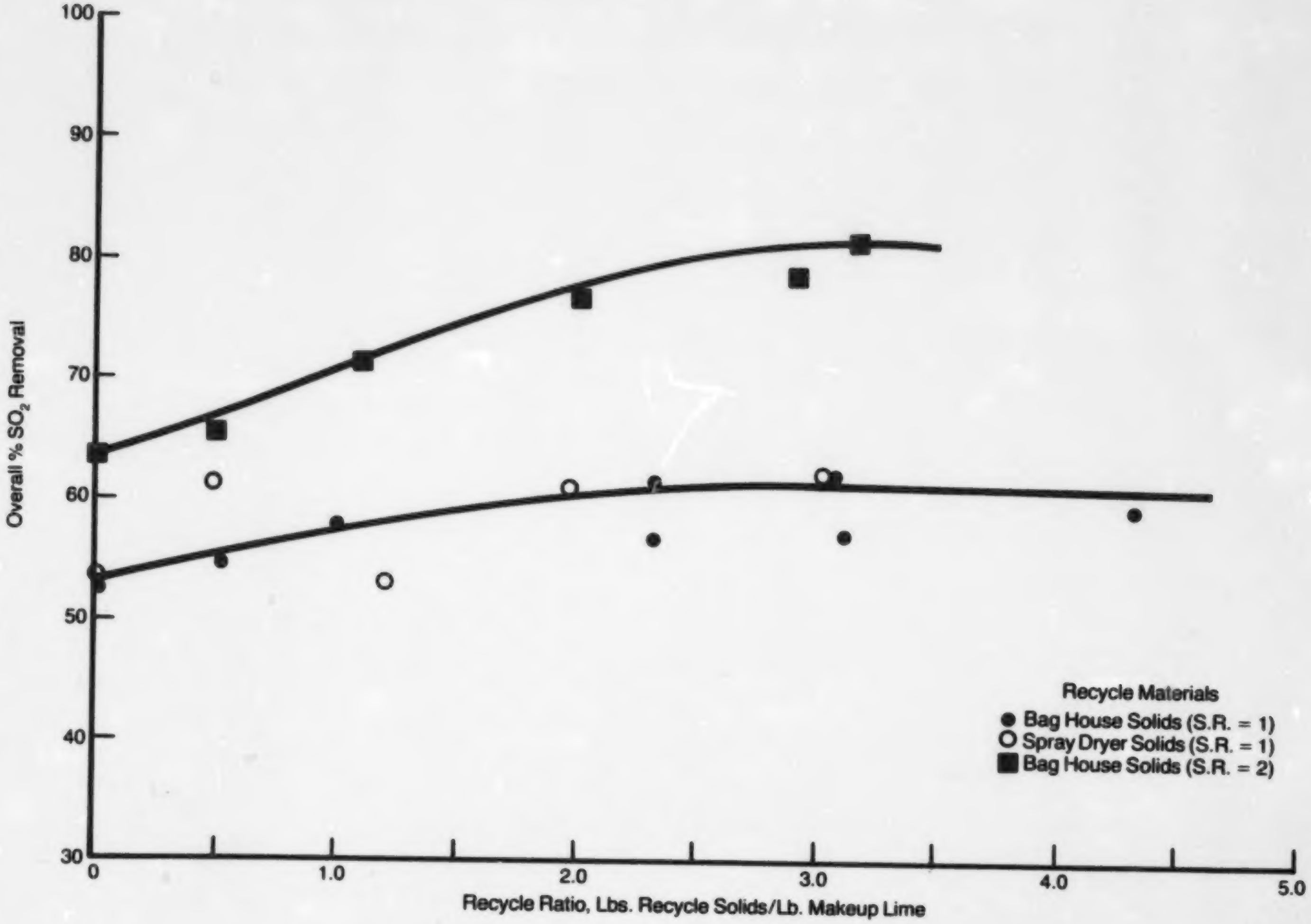

Figure 7 Effect of Recycle Ratio on Overall SO_2 Removal.

FABRIC FILTER PILOT RESULTS

Analysis of fabric filter external variables indicated the following parameters are significant:

- Stoichiometric Ratio (Strong)
- Temperature Approach to Adiabatic Saturation (Strong)
- Fabric Filter Inlet SO_2 Concentration (Weak)
- Air-to-Cloth Ratio (Minimal)

Pressure drop across the fabric filter bags was not considered directly as a variable in the present analysis. Since the pressure drop, which reflects filter cake thickness, was observed to affect SO_2 removal, an appropriate average value during a run was used to analyze the test results.

Stoichiometric Ratio

Percent SO_2 removal in the fabric filter correlates well with fresh lime stoichiometry fed to the spray dryer. SO_2 removal increases directly with increasing fresh lime stoichiometry, as Figure 8 shows. The spray dryer stoichiometry reflects the stoichiometry, based on unreacted lime and SO_2 from the spray dryer, fed to the fabric filter. SO_2 removal efficiency in the fabric filter is based on the inlet and outlet flue gas SO_2 concentrations.

Temperature Approach to Adiabatic Saturation

As observed in the spray dryer, SO_2 removal in the fabric filter is strongly influenced by the flue gas temperature and its proximity to the adiabatic saturation temperature. As the flue gas temperature approaches adiabatic saturation, SO_2 removal efficiency in the fabric filter climbs very rapidly. As Figure 9 shows, SO_2 removal strongly increases when ΔT_{AS} in the fabric filter drops below about 25°F. Higher SO_2 removal near adiabatic

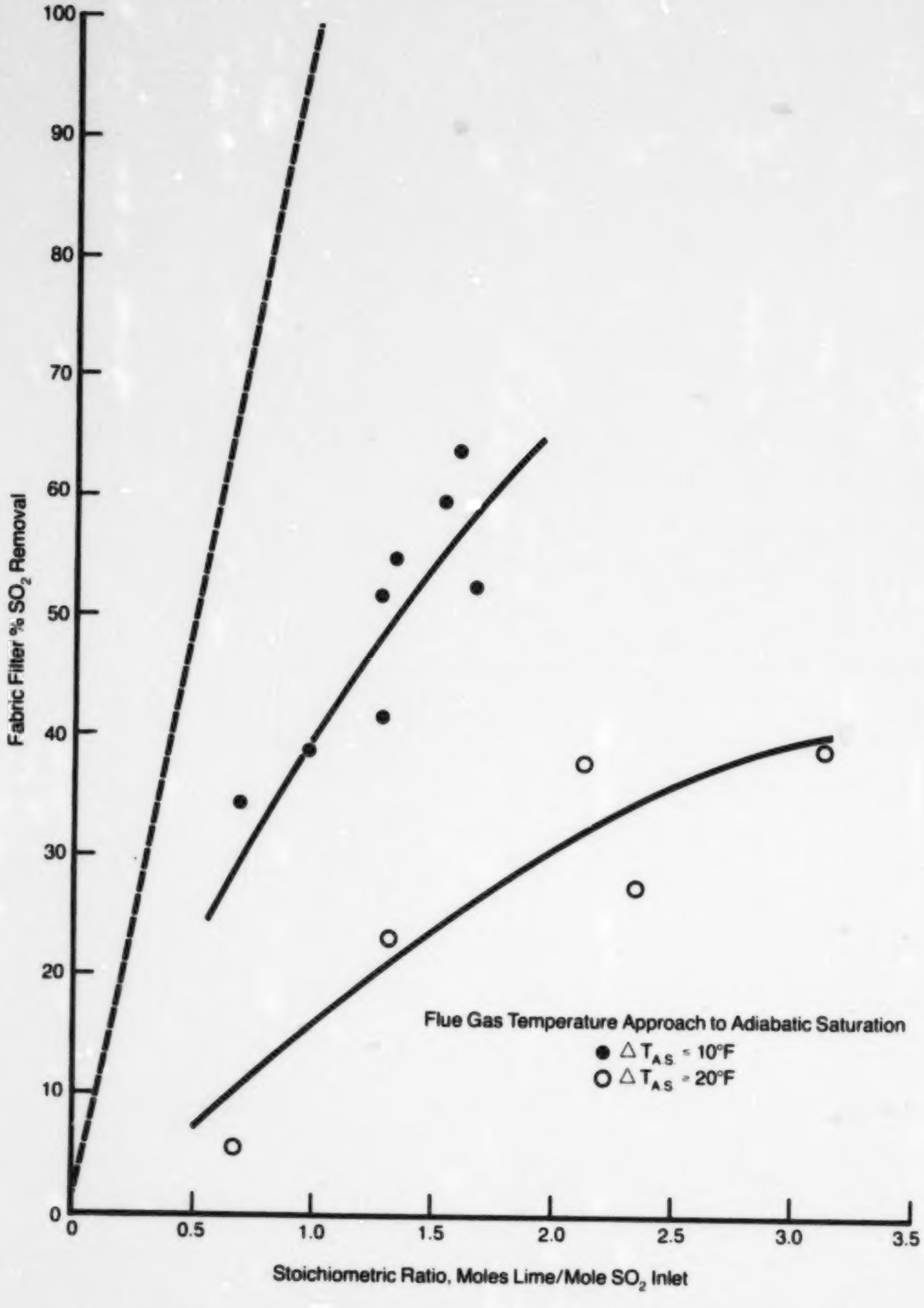

Figure 8 Effect of Stoichiometry on Fabric Filter SO_2 Removal.

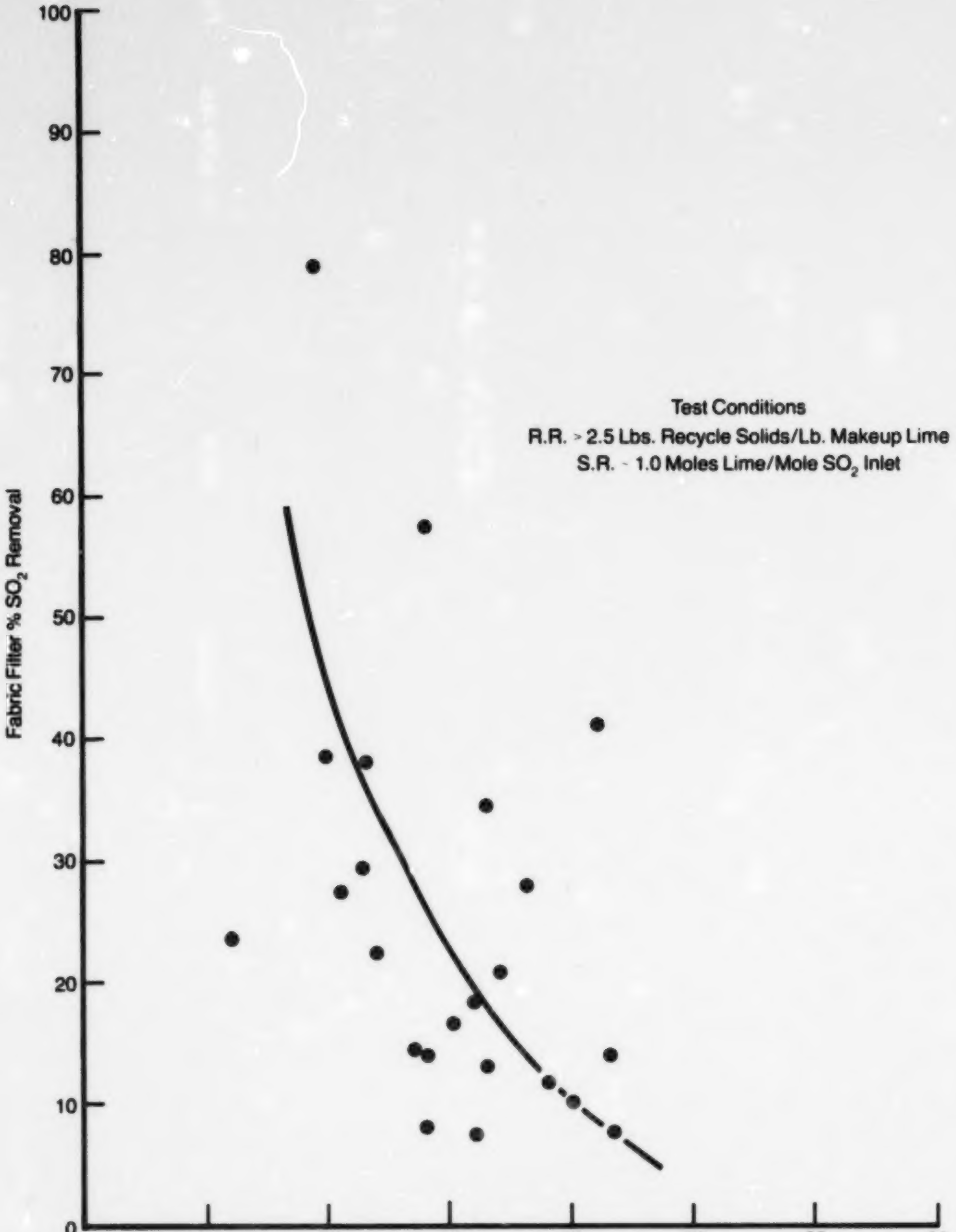

Figure 9 Effect of Flue Gas Temperature Approach to Adiabatic Saturation on Fabric Filter SO$_2$ Removal.

saturation apparently is related to the higher moisture content of the reactant solids in the fabric filter.

Inlet SO_2 Concentration

SO_2 removal in the fabric filter is moderately affected by the SO_2 concentration of the entering flue gas. SO_2 removal decreases with increasing inlet SO_2 concentration to the fabric filter. Figure 10 shows the trend of SO_2 removal with inlet SO_2 concentration at various conditions.

Air-to-Cloth Ratio

Air-to-cloth ratio was found to have a weak effect on SO_2 removal in the fabric filter. Although most of the pilot tests were performed at an air-to-cloth ratio in the vicinity of 2.0 ft/min., the parameter was studied over the range between 1.3 and 2.5 ft/min.

Pressure Drop

SO_2 removal in the fabric filter increases linearly as the pressure drop builds up during operation between cleaning cycles. As mentioned earlier, fabric filter pressure drop has not been utilized as an explicit variable in regression analysis performed to date. The dynamic, or unsteady-state, nature of pressure drop during the course of a test run at otherwise continuous, steady-state conditions very likely has contributed to the greater degree of data scatter of fabric filter compared to spray dryer results. Planned future analyses will examine the data in more detail to delineate better the effect of fabric filter pressure drop on SO_2 removal.

The pressure drop at the beginning of an operating cycle appears to influence the percent SO_2 removal indicating that a thicker initial filter cake improves percent SO_2 removal. Pressure

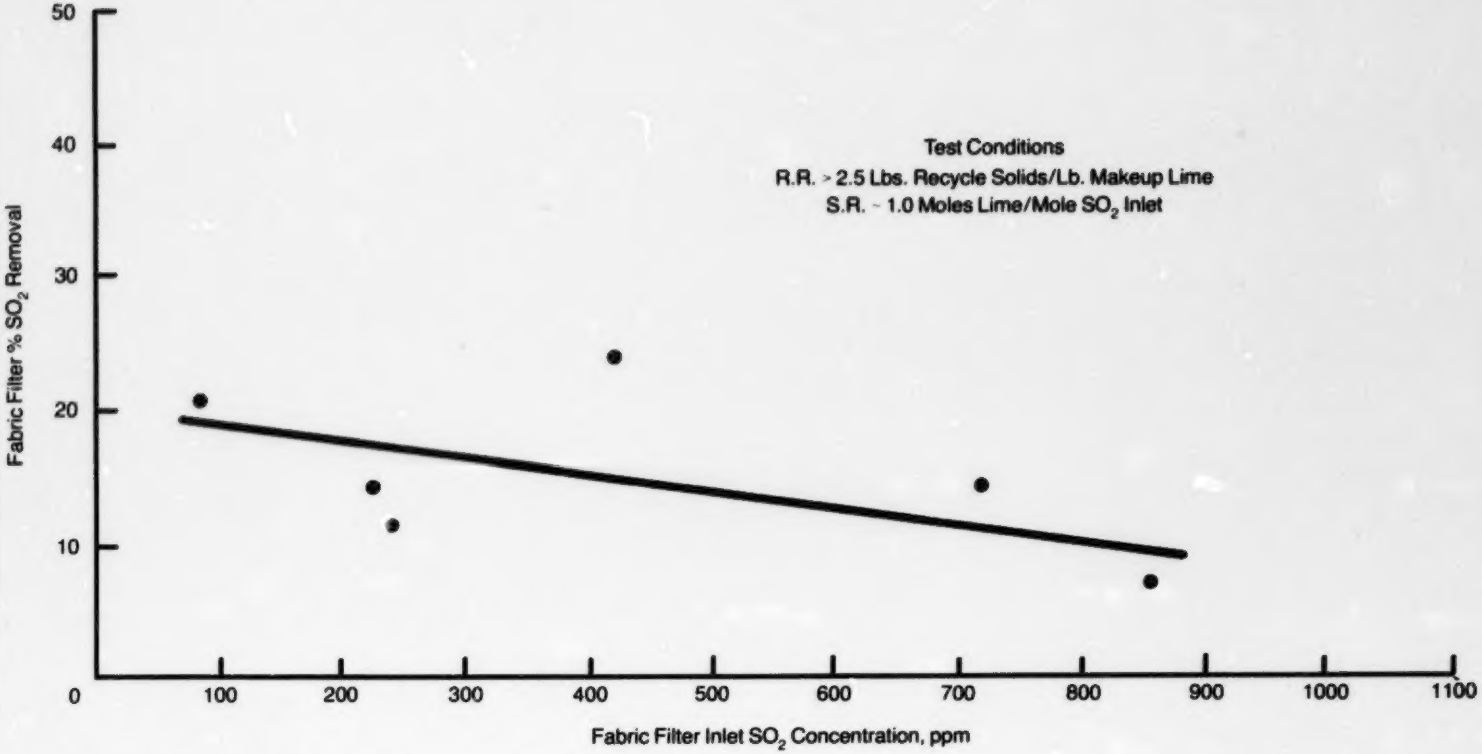

Figure 10 Effect of Fabric Filter Inlet SO_2 Concentration on Fabric Filter SO_2 Removal.

drops over the range of 1-7 in. H_2O were observed during the pilot test program.

Preliminary evidence shows that the pressure increase during an operating cycle occurs more slowly with a mixture of FGD solids and flyash than with flyash alone. This apparent decrease in the specific resistance coefficient is likely due to the presence of larger size FGD solids particles along with flyash in the filter cake.

CONCLUSIONS

The significant external process variables related to SO_2 removal in the spray dryer/fabric filter system have been identified. Lime stoichiometry and flue gas temperature approach to the adiabatic saturation temperature especially affect SO_2 removal in both the spray dryer and fabric filter. Refinements to the present work are in progress and additional parameters that influence SO_2 removal are under investigation. Results of this work and similar efforts by others as well as more basic studies of dry SO_2 scrubbing soon will become available. As a result, the cost, effectiveness and range of applicabilty of this new technology can be refined further. This flow of new information will provide utility and industrial decision makers with a continually improving basis to evaluate dry SO_2 scrubbing for their specific site.

ACKNOWLEDGEMENTS

We thank the host utilities, Public Service of Colorado and Texas Utilities, and their employees for the help and cooperation in enabling Research-Cottrell/Komline-Sanderson to conduct the pilot test programs. We also wish to thank Dr. Theodore G. Brna, EPA Project Officer, for his aid and direction in conducting the EPA-funded portion of the work reported in this paper.

SO_2 REMOVAL BY DRY FGD

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SO_2 REMOVAL BY DRY FGD

ABSTRACT

This paper describes the removal of SO_2 from boiler flue gases using two dry FGD technologies: spray absorption and dry injection. These two dry FGD methods were each investigated in an EPA-sponsored pilot test program, in cooperation with the City of Colorado Springs at its Martin Drake Station. Flue gas from an 85 MW pulverized coal-fired boiler at rates up to 34,000 m^3/hr (20,000 acfm) could be handled by the spray absorber while up to 10,000 m^3/hr (6,000 acfm) was available to the dry injection system.

In a comprehensive series of parametric optimization tests just completed by Envirotech and Anhydro A/S, many aspects of the spray absorption equipment/process interface were investigated, including the spray absorber, sorbent preparation and delivery system, fabric filter, controls, and instrumentation. Process feasibility/demonstration tests were conducted using slaked lime and trona as primary sorbents. Recycled process off-product, consisting of fly ash and reaction products, and limestone with adipic acid additive were tested as supplementary alkalis for the reduction of primary sorbent consumption. Absorber inlet temperature and SO_2 concentration, reactant stoichiometry, and approach to saturation at the absorber outlet were varied as process parameters. Test results, with SO_2 removal rates of up to 94 percent, are discussed in relation to published analytical models.

The use of dry injection of pulverized sodium compounds as a means of SO_2 removal in conjunction with a fabric filter was evaluated in a separate test program. The reactants used were: nahcolite, a naturally occurring sodium bicarbonate; crude lake bed trona ore; and trona upgraded by mining methods to 70 percent sodium bicarbonate content. SO_2 removal by reaction with the dry pulverized sorbents was evaluated as a function of reactant stoichiometry, baghouse inlet temperature, and air-to-cloth ratio. Removal performance as a function of stoichiometry is compared to spray absorption results. Only nahcolite produced performance comparable to spray absorption, with SO_2 removals of up to 90 percent.

INTRODUCTION

The 1979 NSPS for SO_2 and particulate matter promote cleanup of flue gases from utility boilers burning even low-sulfur coals, and the increasing application of fabric filters for control of particulate emissions from such boilers has generated considerable technical and commercial interest in dry FGD. In these processes, sulfur oxides are chemically captured in the form of dry sulfite/sulfate particulates by reactions which occur wholly or in part in a fabric filter.

The most commercially promising dry FGD process is spray absorption. In this process, an alkali sorbent liquid is atomized in a spray dryer using flue gas as the drying medium. The FGD reactions occur between the gas and droplets in intimate association with the various stages of drying, which eventually reduce the reaction products to a suspended solid particulate in the gas stream, some of which drops out in the absorber hopper. The remaining spent sorbent and fly ash are removed from the gas stream in the fabric filter, where further FGD reactions occur with unreacted alkali and residual moisture in the filter cake. A variety of calcium-based and sodium-based alkalis have been evaluated, but slaked lime has the greatest commercial importance.

A second dry FGD technology, which offers unique advantages in terms of the low cost and compactness of the required process equipment, involves the dry injection of sodium compounds. When pulverized sodium carbonates and bicarbonates are pneumatically injected into the inlet duct of a fabric filter, a considerable degree of reaction occurs between the sulfur oxides in the gas and the finely divided sodium compounds. Substantial deposits of naturally occurring sodium carbonates and bicarbonates exist in the western U.S. in the form of nahcolite and trona, but development of mines of sufficient capacity to supply the FGD market as well as potential leachability problems with the process off-product are significant obstacles to widespread commercial development.

Buell and Anhydro A/S have completed the second phase of their joint program for spray absorption FGD technology development, which was initiated in August 1978 in Anhydro's Copenhagen laboratories. This was followed by the construction of a commercial demonstration facility at the Martin Drake Station of City of Colorado Springs. An extensive program of parametric performance and process demonstration tests was undertaken at that facility jointly with EPA and the City of Colorado Springs using a 34,000 m^3/hr (20,000 acfm) spray absorption plant. This system was installed in slipstream configuration on the inlet duct of a 680,000 m^3/hr (400,000 acfm) Buell fabric filter emission control system operating on boiler No. 6, an 85 MW pulverized coal-fired steam generator burning Northwest Colorado coal (see Table 1).

THE TECHNICAL BASIS OF SPRAY ABSORPTION

The spray absorption process utilizes an alkali sorbent liquid to chemically capture sulfur oxides present in flue gas by the formation of alkali sulfites and sulfates which are ultimately reduced to a dry particulate to be collected along with the fly ash. The primary element in the process is a spray dryer which contacts the flue gas stream with a fine spray of sorbent slurry droplets in a manner which promotes chemical absorption of SO_2 by the droplets and results in the drying of the spent sorbent to a particulate suspended in the desulfurized gas stream. The other major elements in the process are the sorbent slurry generating system and the baghouse used to remove the suspended solids from the absorber discharge gas stream. Although other reagents are considered to have potential application to the spray absorption process, lime has the greatest commercial importance. The general discussion in this paper therefore emphasizes lime-based spray absorption.

TABLE 1

Boiler No. 6 Coal Analysis Summary for

Martin Drake Municipal Power Plant

Test Dates	5/24/79	5/30/79	5/31/79	6/1/79	6/5/79	6/6/79	6/7/79
Dry Basis							
kJ/kg	28,423	28,151	28,922	28,327	28,894	28,697	28,434
Btu/lb	12,230	12,113	12,445	12,189	12,433	12,348	12,235
Ash %	7.53	8.14	6.29	7.69	5.64	6.93	8.00
Volatile Matter %	40.13	41.093	41.25	40.89	43.00	43.13	38.99
Fixed Carbon %	52.34	50.68	52.46	51.43	51.36	49.84	53.01
Carbon %	70.31	69.88	71.40	69.88	71.75	70.74	69.85
Hydrogen %	4.96	5.18	5.26	5.00	5.13	5.40	5.15
Nitrogen %	1.58	1.50	1.52	1.59	1.57	1.53	1.52
Sulfur %	0.46	0.52	0.46	0.55	0.52	0.61	0.61
Oxygen %	15.16	14.78	14.98	15.31	15.39	14.79	14.87
As Received							
kJ/kg	24,632	24,353	24,853	24,488	24,799	24,932	24,881
Btu/lb	10,599	10,479	10,694	10,537	10,671	10,728	10,706
Moisture %	13.34	13.48	14.07	13.55	14.17	13.12	12.49
Ash %	6.523	7.05	5.40	6.65	4.87	6.02	7.00
Sulfur %	0.402	0.442	0.40	0.475	0.44	0.53	0.54

Baseline Technologies

The spray absorption process is a new FGD technology which enjoys the great advantage of having little in the way of "barrier technology" problems to overcome. What is required, rather, is a sharpening and fine-tuning of proven technology to make it adaptable to a new and more demanding application. Collection of fly ash in a fabric filter, lime slaking, reaction of slaked lime slurry with SO_2, and spray drying are all relatively mature technologies in their own right, and the refinements necessary to bring these components together into a successful spray absorption system are clearly evolutionary rather than revolutionary in nature.

Spray Drying

The centerpiece of any spray absorption system is the spray dryer, a solids drying device widely used in food processing, dye and chemical, and mineral processing industries. The utility industry represents a new application and places such demands as low pressure drop, low leakage, and low maintenance on the spray dryer, which are usually only of secondary concern to the traditional spray dryer purchaser, who is mainly concerned with the physical characteristics of the off-product and the attainable production rate.

Collection of Fly Ash in a Fabric Filter

The spray absorption process places the additional demands on the baghouse of lower temperature and higher moisture and grain loading at the inlet than in the traditional fly ash application. These contingencies must be met by closer control of baghouse temperature drop, fail-safe temperature controls on the absorber outlet temperature, and reheat of the absorber outlet gas stream in some cases. All of these refinements may be economically accomplished within the scope of existing technology. In over 1000 hours of operation in conjunction with the spray absorber at the Colorado Springs prototype plant and despite numerous system upsets, the test reverse air fabric filter has experienced no bag failures, blinding, or corrosion problems.

Reaction of Slaked Lime with SO_2

Slaked lime is a widely used alkali in water treatment, process pH control, and mineral processing. Lime handling, storage, and slaking equipment is readily available on an "off the shelf" basis that will meet the requirements of the spray absorption process. Just such a standard commercial unit has been used with excellent results at the Colorado Springs prototype plant. The chemical reaction of lime slurry with SO_2 in a spray absorber is similar in most respects to the reaction that occurs in a wet lime scrubber, although the kinetics of the process are somewhat more involved due to the tie-in with the drying process occurring simultaneously, which will be discussed in more detail later.

The Psychrometry of Spray Absorption

The focal point of the spray absorption process is the absorber itself, a spray dryer cast in the dual role of contacting the alkali slurry feed with the flue gas in a manner that promotes chemical capture of SO_2 present in the flue gas by the sorbent, as well as the more normal function of drying the reaction products to a solid particulate material. Since the FGD reactions of primary interest occur only in an aqueous medium, the moisture content of the sorbent is of crucial importance at all stages of the process. Thus from both a standpoint of technical heritage and key process parameters, the spray absorption process must be considered in the context of the classical psychrometry of solids drying.

Application of the Psychrometric Chart

In spray drying, heat and mass transfer are accomplished simultaneously by direct contact of the hot gas and dispersed droplets; the droplets are dried as water vapor evolves from the surface, due to the liquid being heated to a temperature at which its vapor pressure exceeds the partial pressure of the vapor in the surrounding gas. When drying is complete, a substantial portion of the dried product drops into the hopper in the bottom of the chamber. The remaining dried product and evolved vapor are purged from the chamber by the flow of cooled gas, and the product is separated from the gas stream by a particulate collector. For the purposes of this discussion, the psychrometric properties of flue gas will be considered to be the same as those of air. The changes in temperature and humidity of the gas as it passes through the dryer may be conveniently represented by a psychrometric chart, upon which the following features are defined (see Figure 1):

1. Saturation Line

The saturation pressure of water is a function of temperature only, and defines saturation humidity, H_s, as follows:

$$H_s = 0.6219 \frac{p_s}{P - p_s} \qquad (1)$$

H_s = absolute humidity at saturation, kg water/kg dry gas

p_s = saturation pressure of water, mm Hg absolute

P = total pressure of system, mm Hg absolute

At saturation, the vapor pressure of water and the partial pressure of water vapor are both equal to saturation pressure p_s. The curve of H_s versus temperature is called the saturation line.

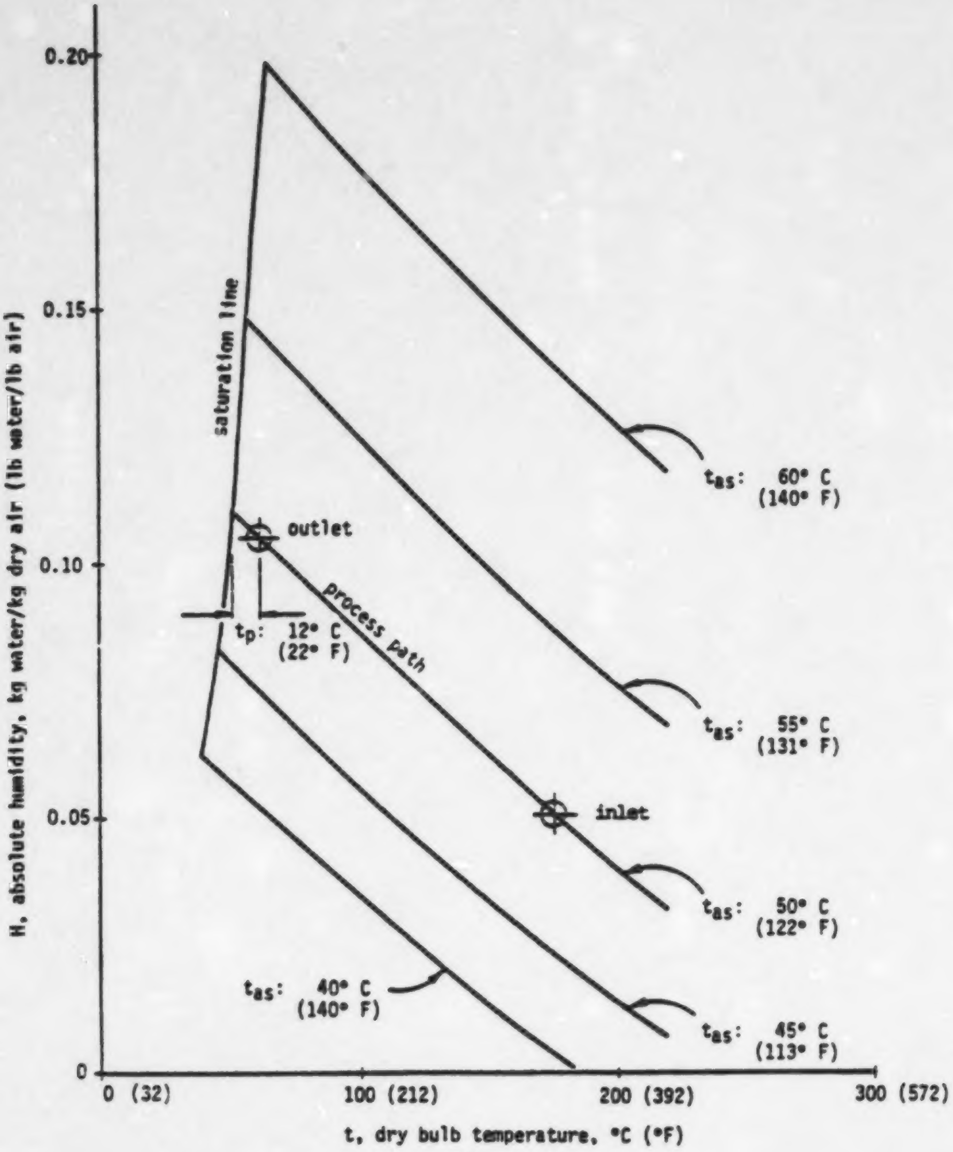


Figure 1 Psychrometric chart for air and water vapor at P = 609.6 mm Hg (24 in. Hg) illustrating typical spray absorption process path.

2. Dew Point

The temperature at which a given line of constant absolute humidity crosses the saturation line is called the dew point, and the locus of dew points is identical to the saturation line. For a given saturation line and humidity, there is always a unique dew point.

3. Humid Heat

The humid heat of moist gas (air) is given by

$$C_s = 0.24 + 0.446 H \qquad (2)$$

where

H = absolute humidity, kg water/kg dry gas

C_s = humid heat of moist gas, kcal/kg-°C

0.24 = specific heat of dry gas, kcal/kg-°C

0.446 = specific heat of water vapor, kcal/kg-°C

4. Adiabatic Saturation Lines

The operation of a well-insulated spray dryer may be modeled fairly accurately as the cooling of a gas stream by the adiabatic evaporation of water, whereby an end point with humidity H_s and temperature t_{as} will be reached if the process continues to saturation. The relationship between temperature and humidity for such a process is represented on adiabatic saturation lines on the psychrometric chart, which are based on the following relationship:

$$H_s - H = \frac{C_s}{\lambda}(t - t_{as}) \qquad (3)$$

where

t = dry bulb temperature, °C

t_{as} = adiabatic saturation temperature, °C

λ = latent heat of vaporization at t_{as}, kcal/kg

5. Wet Bulb Temperature Lines

The wet bulb temperature is established by a dynamic energy balance between water evaporation from and convective heat transfer to a small body such as the wet bulb of a thermometer which is immersed in a gas stream sufficiently large that no measurable change in the humidity or temperature of the gas stream occurs. The relationship between temperature and humidity is expressed as:

$$h_c\ (t - t_w) = k_g'\ \lambda\ (H_w - H_a) \tag{4}$$

where

h_c = average heat transfer coefficient, kcal/hr-m^2-°C

t = dry bulb temperature, °C

t_w = wet bulb temperature, °C

H_w = saturation humidity at t_w, kg water/kg dry gas

H_a = humidity of surrounding air, kg water/kg dry gas

k_g' = mass transfer coefficient, kg/hr-m^2

It happens, that for mixtures of flue gas (air) and water vapor, that $C_s = h_c/k_g'$, making wet bulb temperature lines and adiabatic saturation lines coincident for all temperatures; whereas, in general, they intersect only on the saturation line for other solvents and gases.

6. Approach to Saturation at Outlet

The end point of each adiabatic saturation line is a dew point on the saturation line. This dew point is never reached in dry FGD practice, however, because the spray absorber outlet dry bulb temperature is higher than the saturation temperature, as shown in Figure 1.

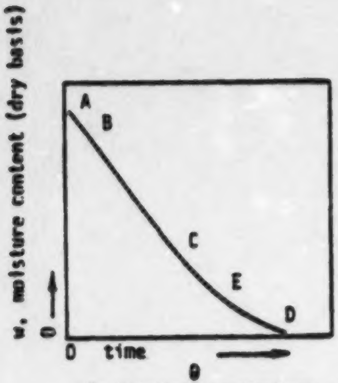
The approach to saturation temperature at the outlet, t_p, where

$$t_p = t_{outlet} - t_{as} \tag{5}$$

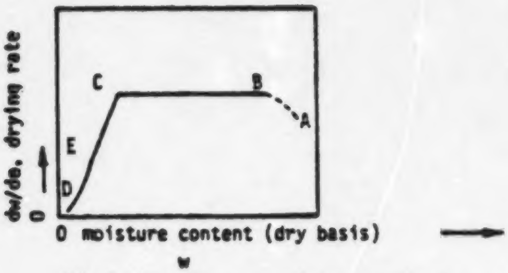
is an important parameter in spray drying and will later be shown to be of crucial importance in spray absorption. The dew point which may actually be reached in practice by cooling the gas downstream of the absorber is defined by the intersection of the absorber outlet humidity line with the saturation line. At the low values of t_p typically seen in spray absorber operation, the outlet dew point so defined will be one or two degrees lower than the adiabatic saturation temperature.

The Effect of Pressure

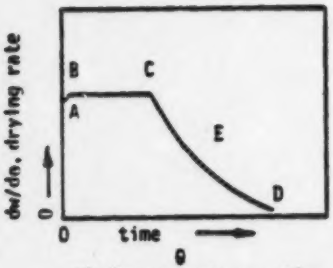
Although there is a unique saturation pressure for a given temperature, the saturation humidity given in Equation 1, and hence the adiabatic saturation lines, will vary according to the total pressure of the system. Thus, for pressures significantly different from sea level barometric pressure, for which standard psychrometric charts are prepared, a special psychrometric chart must be constructed to give accurate prediction of spray absorber performance. For example, the psychrometric chart shown in Figure 1 was especially constructed for the 1830-m (6000-ft) elevation at the Colorado Springs test site, where use



A) Moisture content vs. time



B) Drying rate vs. moisture content



C) Drying rate vs. time

Figure 2 Stages of spray drying: drying rate versus time and moisture content.

the diffusion of moisture away from the evaporating surface, although internal moisture movement becomes increasingly important as the plane of evaporation retreats below the surface of the solid. This condition is represented by point E in Figure 2 and marks the beginning of the second falling rate period, wherein the rate of evaporation is governed by internal moisture movement due to diffusion and capillarity. By the conclusion of the second falling rate period, the dried material will essentially be in equilibrium with the surrounding gas and will pass out of the dryer and be separated from the gas stream as a "dry" powder, whose moisture content will chiefly depend on t_p.

Spray Absorption FGD Chemistry and Reaction Model - Lime Process FGD Chemistry

The sorbent chemical of primary commercial importance is slaked lime. The lime-consuming chemical eactions have been organized by Getler et al.[1] into the following steps:

1. Diffusion of SO_2 and CO_2

$$\begin{aligned} &SO_2\ (g) \rightleftharpoons SO_2\ (aq) \\ &CO_2\ (g) \rightleftharpoons CO_2\ (aq) \end{aligned} \tag{7}$$

2. Dissolution of SO_2 and CO_2

$$\begin{aligned} &SO_2 + H_2O \rightleftharpoons H_2SO_3 \\ &CO_2 + H_2O \rightleftharpoons H_2CO_3 \end{aligned} \tag{8}$$

3. Dissociation in alkaline medium

$$\begin{aligned} &H_2SO_3 \rightleftharpoons H^+ + HSO_3^- \rightleftharpoons 2H^+ + SO_3^= \\ &H_2CO_3 \rightleftharpoons H^+ + HCO_3^- \rightleftharpoons 2H^+ + CO_3^= \\ &SO_2 + H_2O + SO_3^= \rightleftharpoons 2HSO_3- \\ &SO_2 + HCO_3^- \rightleftharpoons HSO_3^- + CO_2 \\ &2SO_2 + H_2O + CO_3^= \rightleftharpoons 2HSO_3^- + CO_2 \end{aligned} \tag{9}$$

4. Dissolution of solids

$$\begin{aligned} &CaCO_3 + H^+ + HCO_3^- \rightleftharpoons Ca^{++} + 2H^+ + 2CO_3^= \\ &Ca(OH)_2 \rightleftharpoons Ca^{++} + 2OH^- \end{aligned} \tag{10}$$

5. Formation of salts

$$Ca^{++} + SO_3^{=} + 1/2\ H_2O \rightleftharpoons CaSO_3 \cdot 1/2\ H_2O\ (s) \qquad (11)$$

$$Ca^{++} + CO_3^{=} \rightleftharpoons CaCO_3\ (s)$$

$$CaSO_3 \cdot 1/2\ H_2O + 1/2\ O_2 + 3/2\ H_2O \rightleftharpoons CaSO_4 \cdot 2H_2O\ (s)$$

All of the above reactions occur in aqueous medium and result in relatively insoluble solid end products, $CaSO_3 \cdot 1/2\ H_2O$, $CaSO_4 \cdot 2H_2O$, and $CaCO_3$, which precipitate from solution as crystalline encrustations on the surfaces of the sorbent particles, as clearly indicated by SEM photographs published by Getler et al.[1] and Downs et al.[2]

The importance of the "competing" reaction of slaked lime with CO_2 in the spray absorption process was initially neglected by some early investigators, including Buell and Anhydro. Despite the fact that CO_2 partial pressure in flue gas is typically at least two orders of magnitude higher than that of SO_2, it was initially treated as an inert gas because CO_2 has lower solubility and because it forms a weaker acid in solution relative to SO_2. In light of early prototype tests at Colorado Springs, however, the importance of the CO_2 reaction was quickly recognized from analysis of the spray absorption process off-product, which indicated that a substantial portion of the lime not reacted with SO_2 was converted to $CaCO_3$. Subsequent prototype tests at Colorado Springs have indicated that the sorbent "lost" to CO_2 may be recovered by means of partial off-product recycle.

Reaction Model

Because of the dependence of spray absorption FGD reactions on water, the stages of reaction are intimately linked to the classical stages of solids drying represented in Figure 2, the warmup and constant rate period, the two falling rate periods, and downstream of the absorber, the equilibrium moisture content plays a vital role in SO_2 removal occurring across fabric filter.

The first stage of reaction occurs in the absorber vessel during the warmup and constant rate (see Figure 2, segment AC) drying periods, where the drying rate is controlled by heat transfer to the droplets, and water vapor diffuses from the saturated droplet surface. The liquid is saturated with $Ca(OH)_2$ and the pH is high, so the reaction rate is controlled by the diffusion of SO_2 and CO_2 into the droplet, which tends to be impeded by the counter-diffusion of water vapor. Downs et al.[2] devised a mathematical model for SO_2 absorption during the constant rate period based on the assumption that gaseous diffusion was rate-limiting. Later it will be shown that this model matches the Colorado Springs prototype data reasonably well for removal of SO_2 across the absorber alone although, as would be expected, it significantly underpredicts performance across the entire system. In the constant rate period, the factors influencing SO_2 removal are those that affect the diffusion of SO_2 into the droplets. Effective mixing of the droplets and the gas stream, high humidity to lengthen evaporation time, and a high sorbent flow rate atomized finely enough to promote mass transfer but not so fine that too rapid drying occurs will all contribute to increased performance.

The second reaction stage also occurs within the absorber during the two falling rate periods (see Figure 2, segments CE and ED) and is far more complex than the first. Depletion of moisture from the droplet has brought the sorbent particles closer together, and significant dissolution of SO_2 and CO_2 as well as the precipitation of calcium salts from solution has lowered the pH and brought the dissolution of fresh $Ca(OH)_2$ into play as a rate-limiting factor. With the onset of the second falling rate period, mass transfer is restricted to interstices between adjacent sorbent particles, which are progressively clogged with crystalline precipitates of reaction products. Eventually, evaporation virtually ceases and the "dry" agglomerations of sorbent particles leave the absorber with an "equilibrium" moisture content. The factors which promote SO_2 removal for the constant rate period also are beneficial during the following rate period. Since the dissolution of $Ca(OH)_2$ is a rate-limiting factor for the falling rate period, it is helpful to have optimized lime slaking, whereby extremely fine, porous particles of sorbent are present in the sorbent slurry.

The third and final reaction period, termed the "psuedo equilibrium" period by Downs et al.[2], occurs downstream of the absorber in the baghouse filter cake. SO_2 removal occurring in this reaction period across the baghouse has been found to be a strong function of t_p, the approach to saturation at the absorber outlet. Low values of t_p will result in high equilibrium moisture content in the absorber off-product, which tends to promote further FGD reactions. In addition to absorber outlet humidity, the other factors that promote SO_2 removal in the falling rate period will also be helpful in the baghouse.

Downs et al.[2] postulated a second mathematical model describing SO_2 removal during all three reaction periods as a function of sorbent surface area, with other factors held constant. The Colorado Springs data will be compared to this model, and it will be shown later in this paper to be an excellent correlation tool.

FGD Chemistry and Reaction Model - Alternate Sorbents

Trona

The material tested was upgraded pulverized trona ore, 95 percent minus 200 mesh, from Owens Lake, CA, consisting of about 85 percent active material. The complex trona molecule, $NaHCO_3 \cdot Na_2CO_3 \cdot 2H_2O$, is soluble in water, and the resulting solution behaves like a mixture of sodium bicarbonate and soda ash solutions. The FGD reactions occur either in an aqueous medium or directly between gas and solid, and may be summarized as follows:

$$\begin{aligned} &2NaHCO_3 \cdot Na_2CO_3 \cdot 2H_2O \rightleftharpoons 2Na_2CO_3 + CO_2 + 3H_2O \\ &Na_2CO_3 + SO_2 \rightleftharpoons Na_2SO_3 + CO_2 \\ &Na_2SO_3 + 1/2\ O_2 \rightleftharpoons Na_2SO_4 \end{aligned} \tag{12}$$

Although reactions occur much more rapidly in an aqueous medium, the stages of drying of the sorbent droplets are less critical than with lime. This is due to the extremely rapid reaction with sodium carbonate, which is a stronger base than lime, and the potential for gas/solid reactions in the dried off-product.

Limestone

The material tested was 99.5 percent pure pulverized calcium carbonate, Q4 grind. As with lime, FGD reactions occur only in aqueous medium and may be summarized as follows:

$$\begin{aligned} &SO_2\ (g) \rightleftharpoons SO_2\ (aq) \\ &SO_2\ (aq) + H_2O \rightleftharpoons H^+ + HSO_3^- \\ &Ca^{++} + HSO_3^- + 1/2H_2O \rightleftharpoons CaSO_3 \cdot 1/2H_2O + H^+ \\ &CaSO_3 \cdot 2H_2O + 1/2\ O_2 \rightleftharpoons CaSO_4 \cdot 2H_2O \end{aligned} \tag{13}$$

Due to the dependence of the reactions on water, it is expected that the reaction model will tie in to the drying stages of the droplets in a manner very similar to the lime process.

Process Off-Product Recycle

Following atomization in the spray absorber and reaction with the flue gas, the process off-product, consisting of dried reaction products and fly ash, is collected and discharged from either the absorber or baghouse hopper. Depending upon the degree of reactant utilization occurring across the absorber and baghouse and the available alkalinity in the fly ash, varying degrees of benefit in terms of increased reactant utilization can be obtained from recycling a portion of the process off-product to the slurry being atomized in the absorber.

If reactant utilization is low, there will be a high percentage of unreacted alkali ($Ca(OH)_2$ as well as $CaCO_3$) in the off-product which will become available for reaction when the material is re-slurried. The recycle slurry will be further enriched if there is additional alkaline material such as CaO, MgO, K_2O, or Na_2O available in the fly ash.

The key factor with respect to fly ash alkalinity is availability. Alkaline material indicated in an elemental analysis will be of no benefit to the spray absorption process if it is bound up in a high silica/alumina ash which takes the form of glassy spheroids.

Fly ash available alkalinity for spray absorption may be readily determined by titration of a test slurry, which will usually show values significantly lower than would be expected from elemental analysis. For example, Colorado Springs fly ash test slurries usually titrated to about 0.4 to 0.6 percent available alkalinity expressed as $Ca(OH)_2$, whereas typical ash analyses, such as those presented in Table 2, indicate considerably higher total alkaline content. These findings are confirmed by an extensive study conducted by EPRI[3].

Since making additional alkaline material available for reaction with SO_2 is the primary objective of recycling part of the off-product, care must be taken in the design of the recycle system so that the maximum benefit is realized. The pH of the slurry to which the recycle material is added must be sufficiently low that a reasonable amount of the recycle alkali will go into solution. Early tests at Colorado Springs showed that if off-product was added to the lime slurry, which was saturated with $Ca(OH)_2$ at a pH of 12, no measurable benefit resulted, regardless of the amount of process off-product added. When a separate slurry of process off-product and water alone was prepared in a separate tank and mixed with the lime slurry at the point of injection to the absorber, however, considerable benefit resulted.

Even if the fly ash has little in the way of available alkalinity, it is plausible that beneficial effects will result from its presence in the atomized slurry droplets by its action as a surface catalyst. The fly ash particles are considerably (approximately 10 times) larger than the elementary particles of slaked lime, and the presence of large inert bodies in the slurry droplet will increase the ratio of surface area to water volume of the droplet as well as provide sites for the formation of crystalline reaction products which will tend to keep the slaked lime particle surfaces "open" to mass transfer for a longer period.

Critical Equipment/Process Interfaces

Spray Absorber

The spray absorber must be carefully optimized so that the sorbent slurry is atomized, contacted with the flue gas, and dried in a manner that promotes maximum capture of SO_2, minimum reactant consumption, and low energy use while maintaining stable and reliable plant operation.

For the dry FGD application, Anhydro uses a centrifugal atomizer, in which the slurry is introduced to a central cavity in a high speed rotating disk called the atomizer wheel and is induced by inertial forces to flow out through radial passages in the wheel. The magnitude of the centrifugal force on the slurry increases as it moves outward in the passage and causes it to be spun into a thin filament of liquid along the passage wall. When the liquid filament reaches the end of the passage at the wheel rim, it stretches out briefly into the surrounding gas and then breaks off to form a spherical droplet whose size is chiefly governed by the viscosity and surface tension of the liquid and

TABLE 2

Fly Ash Analysis for Boiler No. 6

Martin Drake Municipal Power Plant

Constituent	Percentage by Weight		
	Sample 1	Sample 2	Sample 3
SiO_2	65.96	61.38	63.09
Al_2O_3	21.95	24.04	24.02
TiO_2	0.51	0.77	0.62
Fe_2O_3	2.89	4.20	4.16
CaO	2.78	4.68	3.96
MgO	0.60	1.05	0.70
K_2O	1.44	1.69	1.20
Na_2O	0.89	1.27	0.62
P_2O_5	0.59	0.90	0.95
SO_3	0.09	0.14	0.15

the atomizer wheel tip speed. Liquids with low viscosity and surface tension tend to form smaller droplets, and droplet size for a given liquid is inversely proportional to atomizer wheel tip speed. As the droplets move away from the wheel and disperse into the gas stream exiting the gas disperser nozzle, they form an umbrella-shaped spray pattern that is symmetric about the wheel axis, which serves as the zone of initial contact between the sorbent and flue gas. The centrifugal atomizer generates a uniform spray of fine (50 to 80 μm) droplets over a wide range of feed rates without the use of high pressure liquid feed, small orifices that are prone to clogging, or high velocity air streams which tend to promote erosion. For example, the Model CE-250 (250-mm diameter wheel) in use at the Colorado Springs prototype plant has a minimum diameter of 5 mm for the sorbent feed, which is supplied to the atomizer at a pressure of approximately 700 to 1500 mm (28 to 59 in.) H_2O gauge.

The size of the droplets must be controlled to ensure optimum reaction: small droplets provide a large surface area for mass transfer, but they must not be so small that they will dry out before a suitable degree of reaction with SO_2 has occurred. In this regard, the Anhydro atomizer offers the advantage of a belt drive which makes it possible to change atomizer wheel speed, and thus the droplet size, by simply changing the belts and pulleys, which may be done in the field using only the tools furnished with the atomizer. Because of the abrasive character of the slurries atomized in the spray absorption process, a wheel design is used which features silicon carbide inserts in the slurry passages, which may be repositioned as local wear spots appear.

The absorber gas disperser design must be selected for high SO_2 removal without high pressure drop or fly ash abrasion problems. As a result of optimization tests conducted in their Copenhagen laboratory, Anhydro selected a top-entry vaned scroll-type gas disperser, which discharges an annular vortex flow of flue gas down into the chamber on all sides of the atomizer wheel. The gas disperser is equipped with variable vanes which can be adjusted to obtain optimum mixing of the spray and the flue gas.

The absorber chamber must be properly sized in relation to the gas disperser flow volume to ensure that the slurry droplets will have adequate residence time in the chamber for the various stages of reaction with SO_2 and drying to occur. The chamber size also affects the degree of dropout in the absorber, which should be maximized to cut down the particulate loading into the baghouse.

Critical Equipment/Process Interfaces

Slurry Generator

Calcium oxide, CaO, is a white caustic solid also known as lime, burnt lime, quicklime, or caustic lime, and is the reactant of primary interest in the spray absorption process. The hydration of lime to form calcium hydroxide, also known as hydrated lime or slaked lime, in the presence of excess water by the following exothermic reaction,

$$CaO(s) + H_2O(l) \longrightarrow Ca(OH)_2(s) + 15{,}300 \text{ kcal/kg-mole} \quad (27{,}500 \text{ Btu/lb-mole}) \qquad (14)$$

is called slaking.

When high calcium, soft-burned pebble lime is slaked with clean water at a water/lime ratio of 3 to 4, the lime pebbles rapidly disintegrate in an explosive chain slaking reaction. This produces a slurry of extremely fine (0.5 to 4.0μm) slaked lime particles suspended in water which is ideal for use in the spray absorption process.

The primary element of the slurry generator is the lime slaker, which meters the flow of lime and water into an agitator-equipped tank to generate the slaked lime slurry. The slurry is then diluted with more water and processed to remove inert impurities called grits, which generally consist of uncalcined limestone called "core," fragments of kiln brick, and silica, alumina, and ferric oxide contained in the limestone from which the lime was prepared. Grits are undesirable because they are abrasive and prone to settle out in the slurry piping. Slaking under non-optimum conditions, using hard-burned or dolomitic lime with too high a water/lime ratio, or with poor quality slaking water containing excessive dissolved solids, will result in a less reactive slaked lime slurry. Such a slurry is characterized by larger and less porous slaked lime particles, which will produce low sorbent utilization when used in the spray absorption process.

Buell recommends the use of high calcium (88 to 96 percent CaO) soft-burned pebble lime, calcined in a rotary kiln from limestone of oolitic (fossil shell) origin. This lime should be slaked at a water/lime ratio of 3:1 to 4:1 using water of near potable quality (less than 500 mg/l sulfates, less than 1000 mg/l total dissolved solids). The slaking reaction should produce a temperature rise of at least 40°C (72°F) above the incoming water temperature within 3 minutes. For instance, the slaker in use at the Colorado Springs prototype plant produces a temperature rise of 50 to 70°C (90 to 126°F) when slaking a high calcium (96 percent CaO) soft-burned pebble lime, using plant treated water at a water/lime ratio of 3.5:1 to 4:1.

Recycle System

To obtain the maximum effect from the available alkalinity in the spray absorption process off-product, the recycle slurry must be prepared in a separate tank which is vigorously agitated and combined with the lime slurry just prior to injection to the atomizer. Due to the presence of small lumps in the absorber off-product, Buell recommends that the recycle loop be equipped with a classifier (separator) as shown in Figure 3.

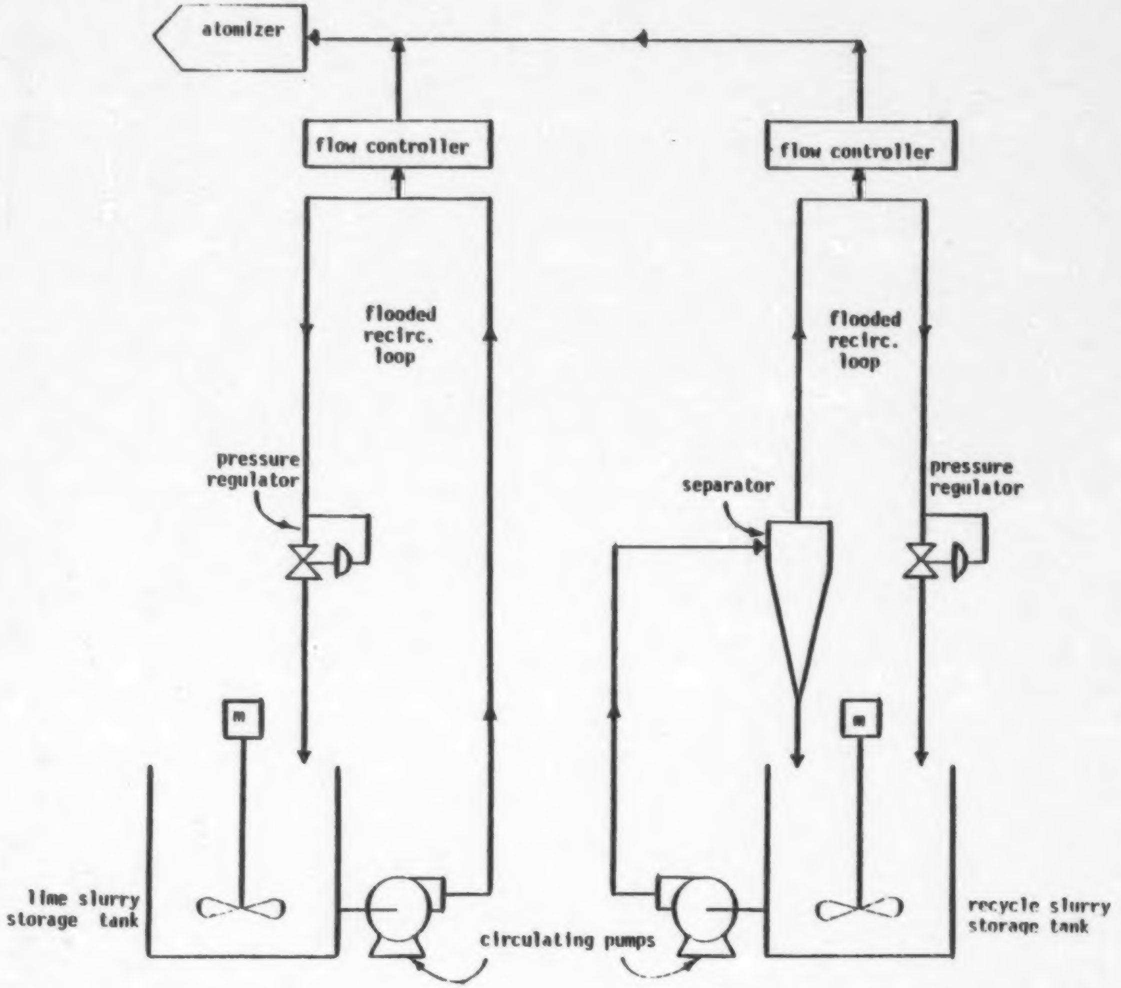

atomizer
flow controller
flow controller
flooded recirc. loop
flooded recirc. loop
pressure regulator
separator
pressure regulator
m
m
lime slurry storage tank
circulating pumps
recycle slurry storage tank

Figure 3 Slurry delivery and control system schematic.

SPRAY ABSORPTION PROTOTYPE TEST PROGRAM

Objectives and Scope

The prototype plant was built and tested to verify the feasibility of spray absorption FGD and demonstrate equipment design integrity under realistic operating conditions. A state-of-the-art system was designed and constructed based on earlier test experience in Anhydro's Copenhagan laboratory, available technical literature, and applicable design practices from related processes.

A detailed test program was planned and executed to cover the objective range of test conditions and operating parameters. Test data was collected, reduced, and correlated to form an applications data base and as a guide to future avenues of investigation.

Equipment Description

The primary elements of the Colorado Springs prototype plant depicted in Figure 4 are a 3.8-m (12.5-ft) diameter spray absorber equipped with a top-entry vaned scroll-type gas disperser, centrifugal atomizer with a 250-mm (10-in.) diameter wheel, a full-scale commercial lime storage, handling, and slaking system, and a reverse air baghouse fitted with full size (30-cm (12-in.) diameter, 9.3-m (30.5-ft) long) fiberglass filter bags. The plant is also equipped with a four-element Buell cyclone collector in parallel with the baghouse at the absorber outlet to provide the capability to independently vary the absorber flow volume and baghouse air/cloth ratio, and a liquid SO_2 storage, vaporizing, and injection system for "spiking" the absorber inlet flow to SO_2 concentrations higher than the 250 to 400 ppm normally present in the flue gas of Martin Drake No. 6 boiler.

Major measurement systems include venturi meters for gas flow measurement at the absorber and baghouse inlets, an extractive SO_2 monitoring system (Dupont Model 460) with probes at the absorber inlet, absorber outlet, and baghouse outlet; a dew point hygrometer with a probe at the baghouse outlet; magnetic flowmeters for the measurement of slurry flows; and extensive pressure and temperature instrumentation. All required equipment and materials for instrument calibration are available on-site, as are chemical laboratory facilities for measurement of slurry reactant concentration, slurry total solids by evaporation, and off-product moisture content.

Plant Operating Parameters

The Colorado Springs prototype plant was originally sized to accommodate the extremely wide range of operating parameters summarized in Table 3. Following the plant startup, shakedown, and initial optimization tests, operation was restricted to a much narrower range of many key parameters.

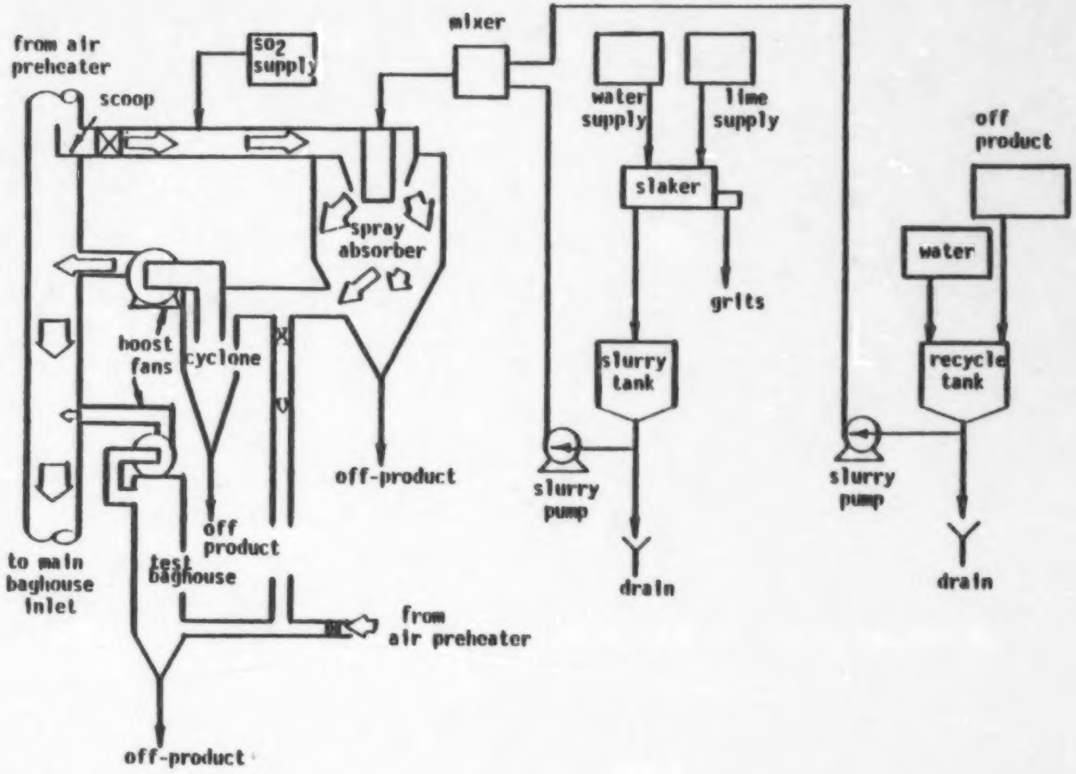

from air preheater
scoop
SO2 supply
mixer
water supply
lime supply
slaker
grits
slurry tank
slurry pump
drain
off product
water
recycle tank
slurry pump
drain
spray absorber
off-product
boost fans
cyclone
off product
test baghouse
to main baghouse inlet
from air preheater
off-product

Figure 4 Spray absorption pilot plant at City of Colorado Springs Martin Drake Power Plant.

TABLE 3

Summary of Operating Parameters for Spray Absorption Prototype Plant in Colorado Springs

	Typical	Minimum	Maximum
ABSORBER			
Inlet flow volume, m^3/hr	14,500	8,500	34,000
(acfm)	(8,500)	(5,000)	(20,000)
Inlet temperature, °C	175	120	200
(°F)	(347)	(248)	(392)
Inlet SO_2 concentration, ppm	1,000	250	2,500
Inlet saturation temperature, °C	48	44	57
(°F)	(118)	(112)	(135)
Outlet temperature, °C	59	51	79
(°F)	(138)	(124)	(175)
t_p, approach temperature, °C	11	7	22
(°F)	(20)	(13)	(40)
BAGHOUSE			
Inlet flow volume, m^3/hr	5,100	1,700	10,200
(acfm)	(3,000)	(1,000)	(6,000)
ATOMIZER			
Wheel speed, rpm	12,500	6,300	14,000
SLURRY SYSTEM			
Lime slurry flow, liters/hr	227	113	680
(gpm)	(1.0)	(.50)	(3.0)
Total atomizer feed, liters/hr	454	227	1,362
(gpm)	(2.0)	(1.0)	(6.0)
Lime slurry total solids, percent	15	8	29

SPRAY ABSORBER OPTIMIZATION TESTS

An early series of tests were dedicated to optimization of the critical equipment/process interfaces involved with the spray absorber. The following parameters were investigated:

Atomizer Wheel Speed

A finely atomized spray of sorbent slurry droplets is desired to provide a large surface area for rapid absorption of SO_2, but with droplets large enough to keep from drying out before a satisfactory degree of reaction has occurred. The size of the slurry droplets is a strong function of atomizer wheel tip speed: droplet diameter is inversely proportional to rpm for a given wheel. Thus there must be an optimum atomizer wheel speed to give the highest FGD efficiency. The results of a test to determine this speed are shown in Figure 5A, where atomizer rpm was varied with other parameters held constant. A plot of comparative SO_2 removal, where the removal percentage for each point is divided by that for the baseline case, indicates that the optimum results were obtained at a wheel speed of 12,500 rpm, which corresponds to a tip speed of 164 m/sec (537 ft/sec).

Absorber Residence Time

The flow of gas through the absorber must be high enough for efficient mixing of the gas and spray but not so high that a proper degree of drying cannot occur in the chamber. The residence time, defined as the chamber volume divided by the absorber outlet gas flow volume, must therefore be controlled to an optimum value so that high SO_2 removal rates are obtained and such problems as chamber wall deposits from too high residence time and wet off-product from too low residence time are avoided in the operation of the spray absorber. The results of a test to determine this optimum residence time are shown in Figure 5B, where inlet flow volume was varied, while all other parameters were held constant. The optimum flow volume of about 14,500 m^3/hr (8500 acfm) corresponds to a residence time of 10 seconds.

Gas Disperser Vane Angle

The secondary swirl vanes of the Anhydro gas disperser can be rotated to obtain the optimum gas flow pattern in the reaction zone where the gas stream mixes with the sorbent spray. The results of a test to determine the optimum angle are shown in Figure 5C, and indicate that an angle between 15 and 30 degrees from the vertical will give optimum performance.

The optimum values for the above parameters have been used in all subsequent testing and are also incorporated into commercial plant designs.

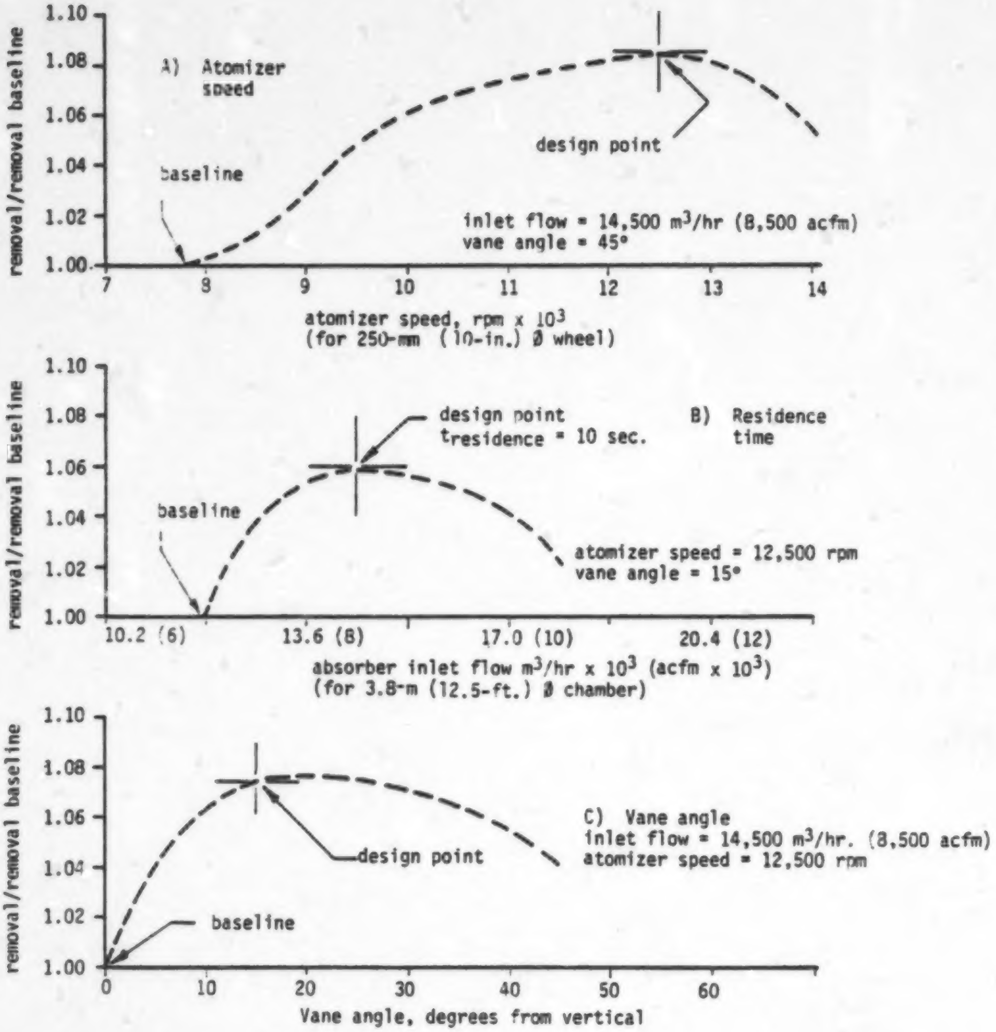


Figure 5 Spray absorber optimization test results comparing SO_2 removal with atomizer speed, residence time, and vane angle at:

Approach to saturation, t_p = 17° C (30° F), stoichiometric ratio = 1.0, inlet temperature = 177° C (350° F), 1000 ppm SO_2.

Particulate Collection Efficiency

A spray absorber of the type used in the Buell/Anhydro spray absorption system is designed for a high degree of product recovery from the chamber hopper. In fact, it is aerodynamically similar to an axial entry cyclone collector. By virtue of its top entry gas disperser and horizontal outlet pipe extending to the chamber centerline at the top of the hopper, the absorber in use at the Colorado Springs prototype plant has a collection rate of about 55 percent of the total solids recovered. A high rate of dust collection in the absorber is extremely desirable because of the relief provided to the baghouse in terms of dust loading.

Pressure Drop

The Anhydro gas disperser's vaned scroll design is a major factor in the relatively low pressure drop across the absorber. During optimization tests, the range of absorber flange-to-flange differential pressure was 41 to 56 mm (1.6 to 2.2 in.) H_2O.

Moisture Carryover

During tests at the Colorado Springs prototype plant, system upsets occurred from time to time wherein wet particulate was discharged from the absorber hopper dump valve. Fear of such material entering the baghouse and causing either blinding or caustic attack on the bags is a real concern of potential operators of spray absorption systems. All observations to date at the prototype plant, however, indicate that no wet material ever leaves the absorber hopper and enters the baghouse during such upsets. Excess moisture tends to cause the normally dusty off-product to agglomerate into much larger soil-like particles which are virtually all removed from the gas stream by the highly efficient cyclonic action of the absorber. Hence the only real potential problem is one of hopper plugging in the absorber, which can cause no permanent damage and allows a lot of time for correcting the condition before the hopper fills. Furthermore, due to the small number of absorber hoppers in a spray absorption system, a good deal of "overkill" in devices to prevent pluggage, such as extensive heat-tracing, air blasters, pneumatic hammers, and the like, can be applied with minimal cost impact.

Advantages Over Alternate Designs

Absorber designs with the gas discharge point at or near the bottom of the hopper do not share the advantages due to high dust dropout enjoyed by the Anhydro design. In a spray absorption system utilizing hopper bottom or near-bottom gas discharge, the baghouse suffers from very high dust loadings (approximately double) because little or no dust is removed upstream in the absorber.

For the same reason, the baghouse has no protection from wet particulate and is "mudded" during a system upset. Wetting of the bags is particularly detrimental in spray absorption due to the likelihood of caustic attack by calcium compounds present in the "mud" in contact with the fiberglass cloth.

Parametric Performance Tests

Results of the parametric performance tests for straight-through lime and partial off-product recycle are shown in Figures 6 through 12.

The crucial role of approach to saturation (t_p) at the absorber outlet is best illustrated by Figure 6, but is apparent throughout the data. The Colorado Springs prototype tests demonstrated the capability of running continuously at t_p = 8°C (15°F) with a dry off-product at 10 second residence time when precise control of the outlet temperature is maintained. Buell considers it prudent, however, to design systems with an 11°C (20°F) minimum t_p retaining any closer approach to saturation as a technical contingency.

The substantial benefit gained through partial recycle of process off-product is shown in Figures 7 and 8. This performance benefit was derived with a fly ash of 0.4 to 0.6 percent available alkalinity and thus must be due to $Ca(OH)_2$ and $CaCO_3$ recovered from the off-product and surface catalyst effects. A greater recycle benefit is realized at t_p = 17°C (30°F) than at 11°C (20°F) because the higher t_p produces lower reactant utilization on a straight-through basis, which leaves more potential reactant in the off-product.

The effect of fabric filter removal is illustrated in Figures 9 through 11 and clearly indicates the strong dependence of baghouse SO_2 removal on t_p. Such a relationship would be anticipated since t_p is the main factor governing the residual moisture in the "dry" powder leaving the absorber, with a lower t_p giving a higher moisture. Thus, much of the performance gain observed at low t_p may be attributed to the increasing role of the "psuedo equilibrium" stage of the FGD reaction model as off-product moisture is increased.

The effect of inlet SO_2 concentration is shown in Figure 7, where relatively constant performance is obtained at 1000 and 1500 ppm but a considerable fall-off in performance is experienced at 2000 and 2500 ppm for straight-through lime. A high degree of performance recovery is possible, however, if partial recycle of the off-product is used because of the relatively poor reactant utilization on a straight-through basis.

The results of reducing the absorber inlet temperature by use of a water-cooled heat exchanger in the slipstream duct are shown in Figure 12 and indicate a neglible effect on performance. Several attempts were also made to pre-cool at the absorber inlet by water sprays, but because of the small 660-mm (26-in.) duct diameter, it was

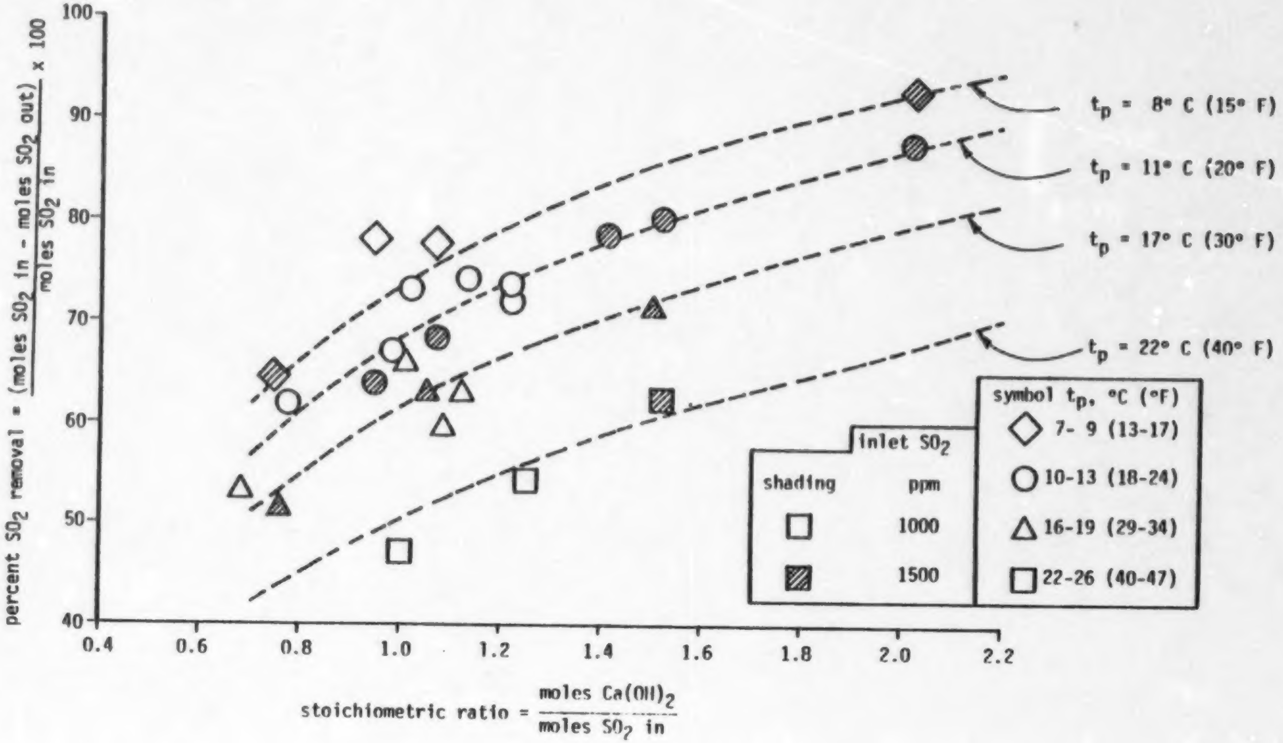


Figure 6 Data reduction for dry FGD tests (straight lime): percent SO_2 removal vs stoichiometric ratio for 1000 & 1500 ppm SO_2 for different values of approach temperatures, t_p, and residence time = 10 sec.

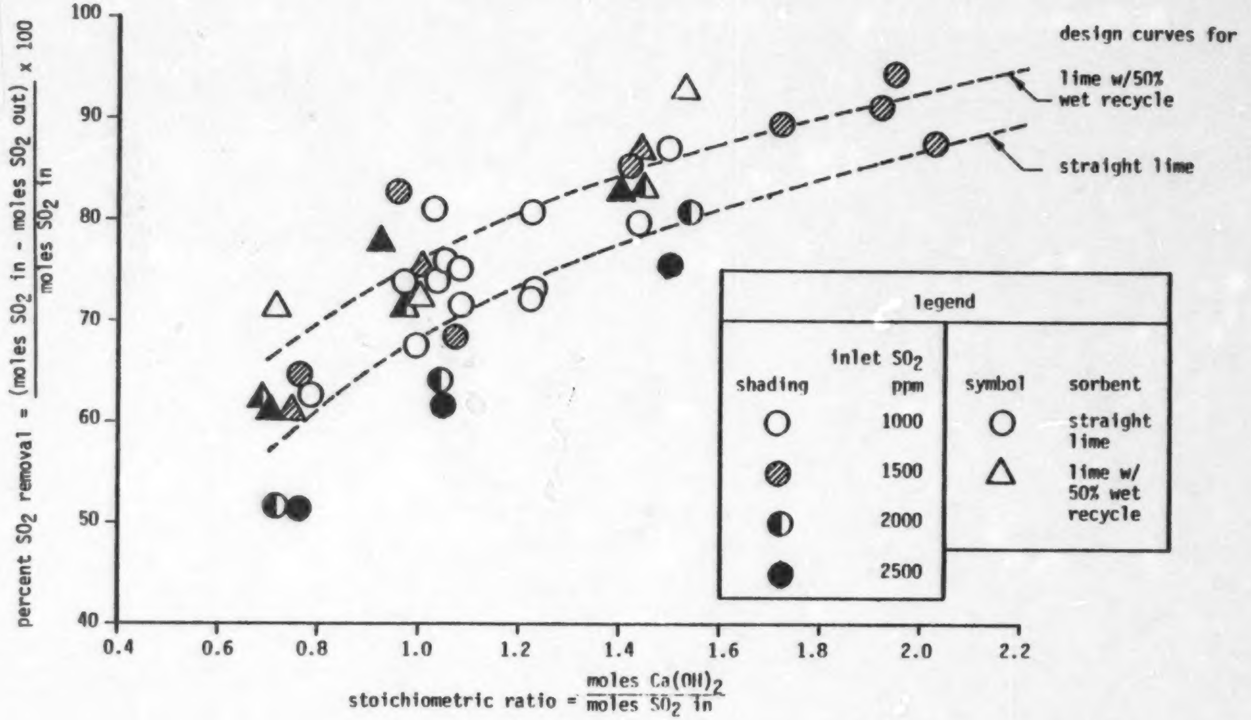


Figure 7 SO_2 removal for dry FGD tests:
percent SO_2 removal vs. stoichiometric ratio
for 10° to 13° C (18° to 24° F) approach to saturation at the outlet, t_p,
inlet gas temperature = 150° to 180° C (290° to 360°F), residence time = 10 sec.

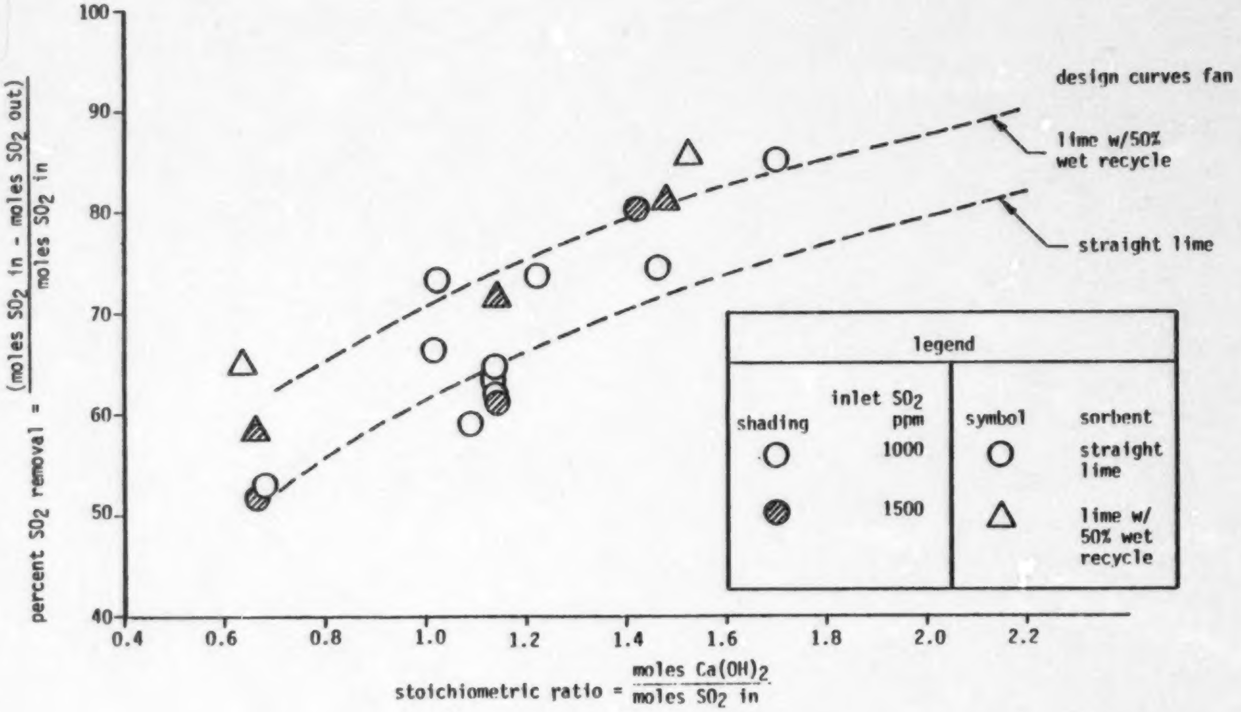


Figure 8 SO_2 removal data reduction for dry FGD tests:
percent SO_2 removal vs. stoichiometric ratio
for 16° to 19° C (29° to 34° F) approach to saturation at the outlet,
t_p, inlet gas temperature = 150° to 180° C (290° to 360° F),
residence time = 10 sec.

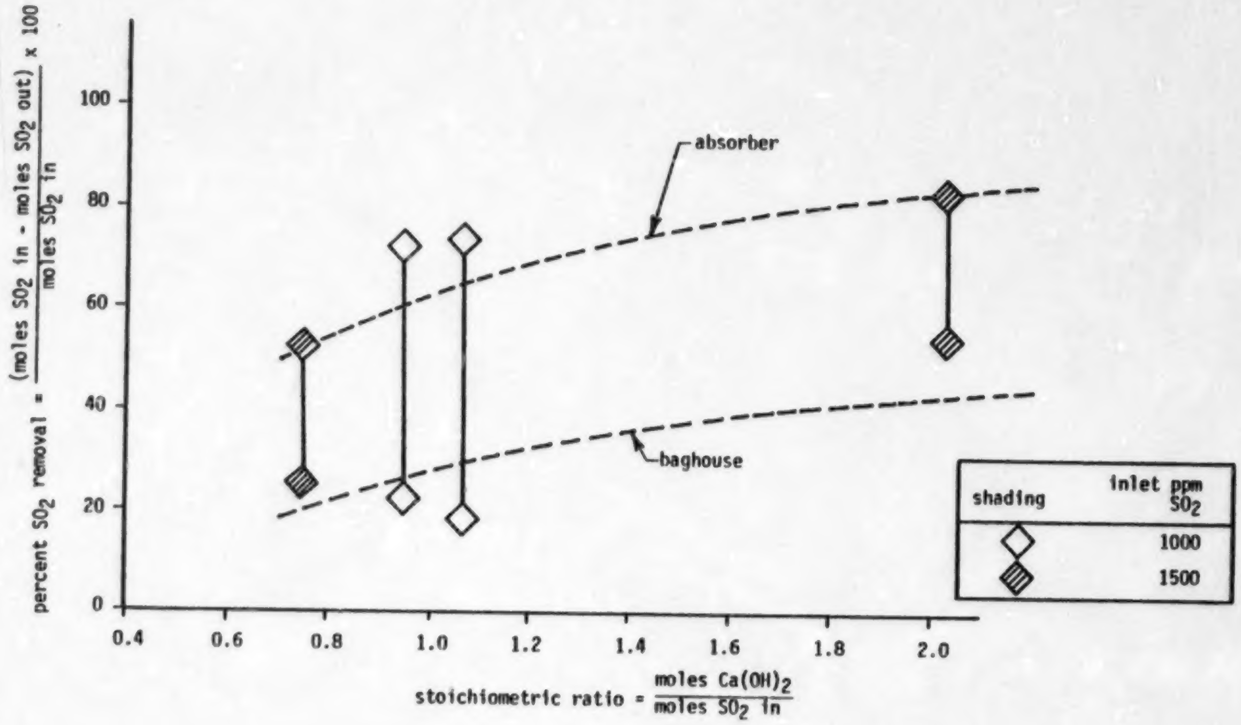


Figure 9 Data from spray absorber prototype plant: percent SO_2 removal vs. stoichiometric ratio for flange-to-flange absorber & baghouse at 1000 & 1500 ppm SO_2 with 7° to 9° C (13° to 17° F) approach to saturation at the outlet, inlet gas temperature = 160° to 180° C (290° to 360° F), and residence time = 10 sec.

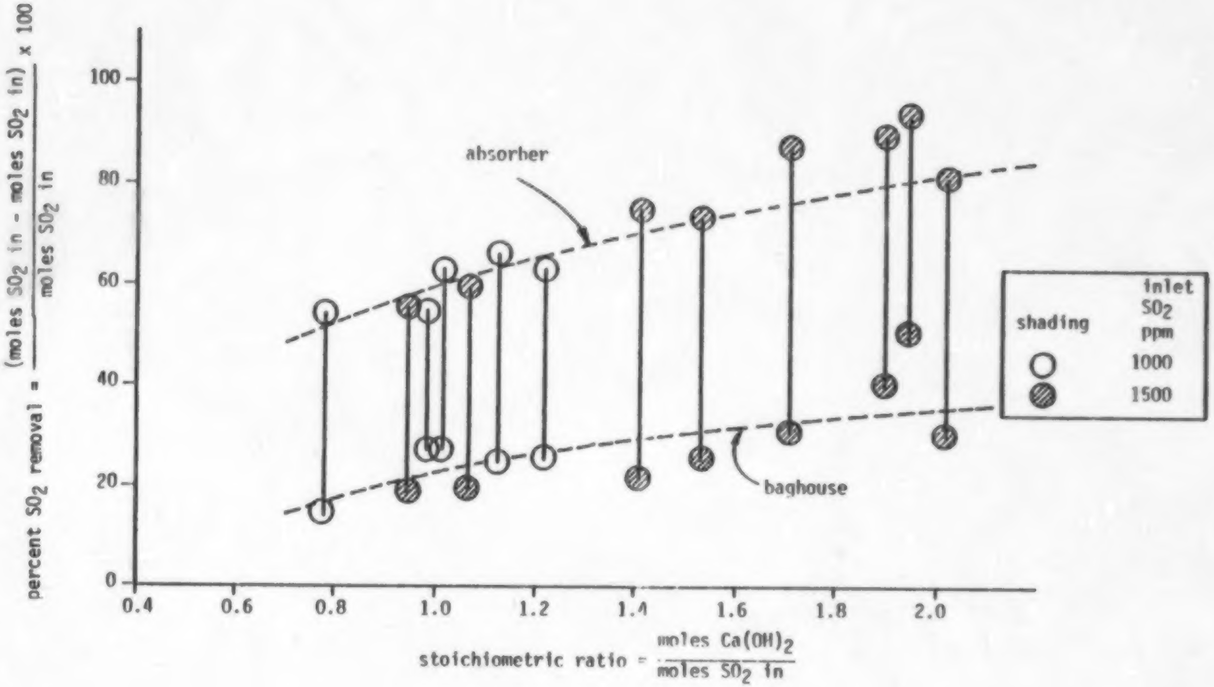


Figure 10 Data from spray absorber prototype plant: percent SO_2 removal vs. stoichiometric ratio for flange to flange absorber and baghouse @ 1000 and 1500 ppm SO_2 with 10° to 13° C (18° to 24° F) approach to saturation at the outlet, inlet gas temperature = 160° to 180° C (290° to 360° F), and residence time = 10 sec.

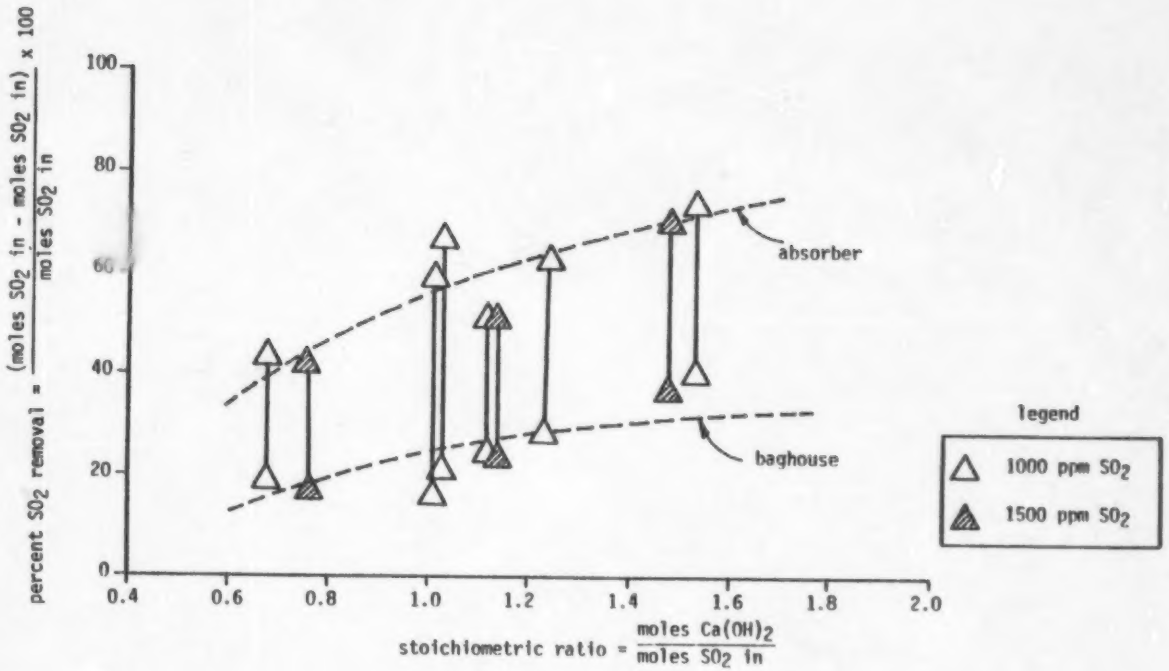


Figure 11 Data from spray absorber prototype plant: Percent SO_2 removal vs. stoichiometric ratio for flange to flange absorber & baghouse @ 1000 & 1500 ppm inlet SO_2, with 16° to 19° C (29° to 34°F) approach to saturation at the outlet, inlet gas temperature = 160° to 180°C (290° to 360°F), and residence time = 10 sec.

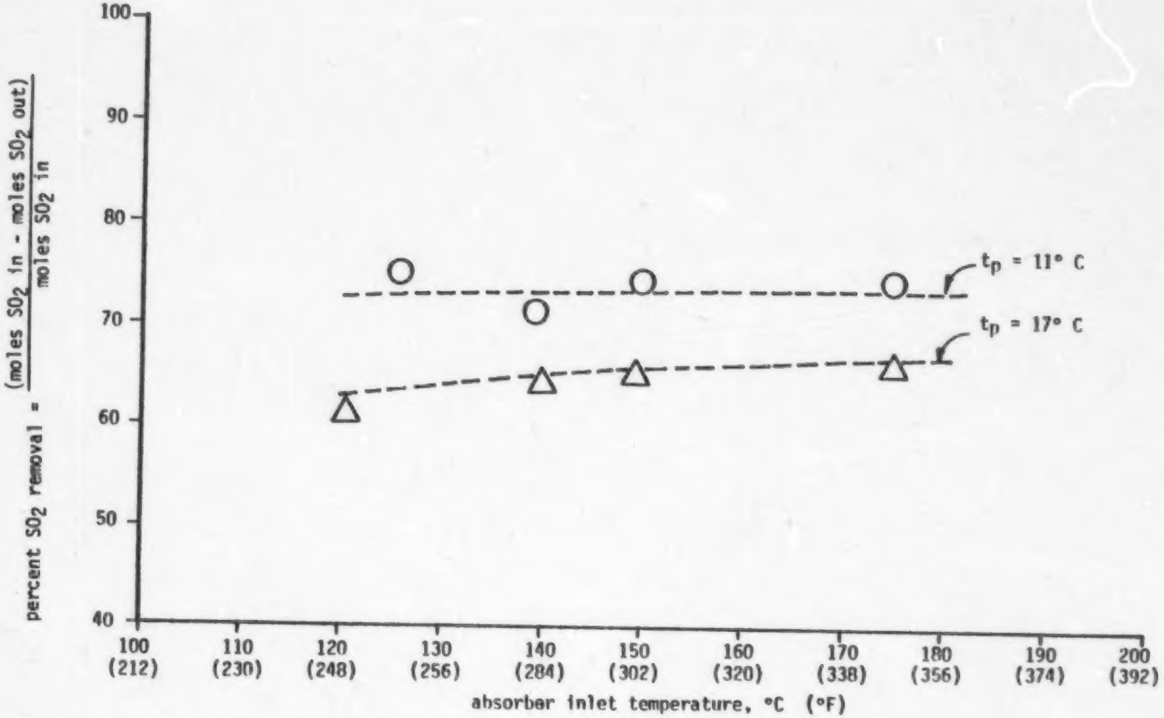


Figure 12 Data from prototype plant - test 11: percent SO_2 removal vs. absorber inlet temperature for 1000 ppm SO_2 inlet, with S.R. ≅ 1.0 at two approaches to saturation at the outlet, t_p, and residence time = 10 sec.

impossible to prevent fouling of the walls even with a 40° spray angle. The one test point run indicated results sufficiently promising that a 1220-mm (48-in.) diameter spray tower was constructed at the prototype plant for tests using a water spray prequench.

Endurance Test

The demonstration of sustained performance is an important part of any process feasibility test program. Such a demonstration was accomplished by the successful completion of a 100-hour endurance test on the Colorado Springs prototype plant. Data taken during that test are tabulated in Table 4, and indicate steady performance despite considerable fluctuations in inlet temperature and SO_2 concentration during the cyclic operation of Martin Drake No. 6 boiler. No equipment failure or performance deterioration was observed, and the plant controls operated satisfactorily in the "automatic" mode.

Utilization of Plant Wastewater

An important concern of many potential buyers of spray absorption FGD systems is the capability of utilizing plant wastewater to meet part or all of the process water requirements. To test this, a typical wastewater was extracted from the feed line to the No. 5 cooling tower at the Martin Drake Station for use in the prototype plant. An analysis of this water, summarized in Table 5, indicates that it had sulfates and total dissolved solids far in excess of Buell's recommended limits for slaking water. These suspicions were confirmed when slaking was performed with this water. Despite high slaking temperatures, a slaked lime slurry of very poor quality resulted, with particles so large that about 30 percent of the lime slaked was removed by the slaker's grits removal system. Large particle size in the sorbent slurry resulted in poor performance, as may be seen from the results of tests 14A, B and E, in Table 6. From the results of tests 14C and D, however, it appears that using cooling tower water for dilution will result in performance equivalent to that obtained using treated water for dilution. Since slaking water accounts for only 10 to 30 percent of water use in spray absorption, it appears that a substantial use of wastewater, such as cooling tower blowdown, may be feasible.

In-Situ Resistivity Tests

The results of in-situ resistivity tests conducted at the absorber discharge are shown in Figure 13. As expected, the resistivity is decreased with higher stoichiometries.

Comparison with Published Analytical Models

Constant Rate Period

Downs et al.[2] proposed a mathematical model based on gaseous diffusion as the rate limiting process for SO_2 absorption during the constant rate

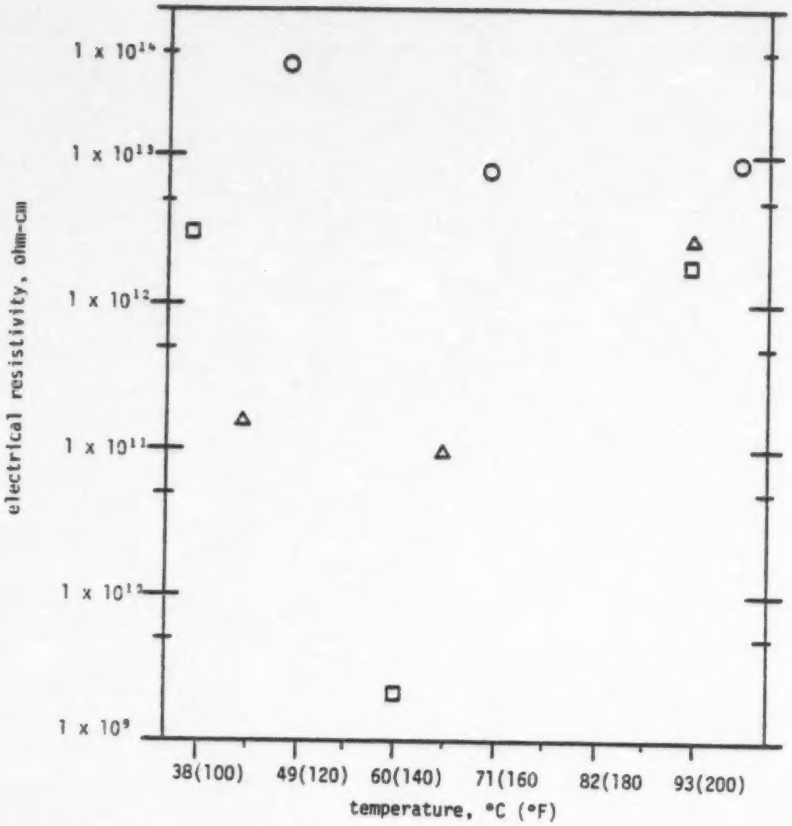


Figure 13 Spray absorption off-product (straight lime) resistivity vs. temperature.

○ sample 1 (15-A) 7-22-80 S.R. = 1.0
△ sample 2 (15-B) 7-23-80 S.R. = 1.2
□ sample 3 (15-C) 7-24-80 S.R. = 1.4

S.R. denotes stoichiometric ratio.

TABLE 4

Process Parameters for 100-hour Endurance Test

Time hrs	Abs[a] Inlet Temp. °C	(°F)	Abs Outlet Temp. °C	(°F)	Adiabatic Sat. Temp. °C	(°F)	SO_2 Concentration (wet) Abs Inlet ppm	Abs Outlet ppm	B.H.[b] Outlet ppm
0	157	(315)	65	(149)	48	(118)	303	132	97
4	159	(318)	64	(147)	47	(117)	307	136	106
8	155	(311)	64	(147)	47	(117)	307	140	114
12	142	(288)	63	(145)	47	(117)	325	134	100
16	139	(282)	62	(144)	45	(113)	260	101	62
20	155	(311)	63	(145)	47	(117)	302	96	57
24	157	(315)	65	(149)	47	(117)	301	155	113
28	158	(316)	64	(147)	47	(117)	343	141	111
32	155	(311)	64	(147)	48	(118)	422	153	126
36	152	(306)	64	(147)	47	(117)	346	107	71
40	139	(282)	61	(142)	46	(115)	286	110	76
44	153	(307)	63	(145)	47	(117)	314	103	55
48	157	(315)	64	(147)	47	(117)	312	119	89
52	150	(302)	63	(145)	48	(118)	337	126	91
56	151	(304)	64	(147)	48	(118)	323	116	106
60	142	(288)	63	(145)	46	(115)	290	107	80
64	138	(280)	62	(144)	46	(115)	261	90	55
68	146	(295)	64	(147)	48	(118)	267	108	65
72	159	(318)	64	(147)	48	(118)	266	118	76
76	163	(325)	65	(149)	48	(118)	252	110	80
80	160	(320)	64	(147)	48	(118)	318	149	121
84	148	(298)	63	(145)	47	(117)	347	128	100
88	141	(286)	62	(144)	46	(115)	318	139	98
92	151	(304)	63	(145)	47	(117)	386	157	115
96	159	(318)	64	(147)	48	(118)	331	140	98
100	157	(315)	64	(147)	48	(118)	362	160	127

[a] Abs denotes spray absorber.

[b] B.H. denotes baghouse.

2.

TABLE 5

Cooling Tower Water Analysis - Martin Drake Station

Tower No. 5 - 7/25/80

pH		6.6
mg/l	total hardness[a]	1044
mg/l	Ca hardness[a]	752
mg/l	Mg hardness[a]	292
mg/l	Silicate	122
mg/l	Sulfate[a]	6348
mg/l	Cl ion[a]	525
mg/l	Na ion[a]	5871
mg/l	Total Phosphate	8.1
mg/l	Ortho Phosphate	5.2
mg/l	Organic Phosphate	2.8
mg/l	Cl - 144	22.1
mg/l	Cl - 68	38.0
mg/l	Chromate	6.0

[a] Expressed as calcium carbonate

TABLE 6

Spray Absorption FGD Performance
Using
Cooling Tower Blowdown Water
(See Table 5 for Analysis)

Test No.	Cooling Tower Water for Slaking?	Dilution?	Approach Temp., t_p °C	(°F)	Stoich. Ratio	Percent SO_2 Removal
14A	Yes	No	17	(30)	1.04	45.9
14B	Yes	No	11	(20)	1.03	53.9
14C	No	Yes	17	(30)	1.46	78.2
14D	No	Yes	11	(20)	.99	61.0
14E	Yes	Yes	17	(30)	1.08	50.1

period, which is valid for high stoichiometries and is written:

$$N_T = 0.62 \ln \frac{(t_{in} - t_{as})}{(t_{out} - t_{as})} \qquad (15)$$

where

N_T = $-\ln (1 - E)$

E = SO_2 removal fraction

t_{in} = absorber inlet temperature, °C

t_{out} = absorber outlet temperature, °C

In Figure 14A, absorber removal test data for stoichiometric ratio (S.R.) of 1.2 or more is compared with this model and a reasonable degree of agreement is indicated. Such a model would, of course, considerably underpredict removal across the entire system.

Overall Removal

Downs et al.[2] have also proposed a correlation method for overall removal performance as a function of stoichiometric ratio with other factors held constant. The data should correlate in the form

$$-\ln (1 - E) = K \cdot S.R. \qquad (16)$$

where

K = correlation constant

S.R. = stoichiometric ratio, moles $Ca(OH)_2$/mole SO_2 in

A plot of overall removal performance for $10 \leq t_p \leq 13°C$ $(18 \leq t_p \leq 24°F)$ is made in the form of the above correlation in Figure 14B, and good agreement with the model is shown.

SPRAY ABSORPTION TEST RESULTS - ALTERNATE SORBENTS

Trona

High SO_2 removal was obtained, with extremely high reactant utilization, using trona as a sorbent, as may be readily determined from the results plotted in Figure 15. Due to the hygroscopic character of sodium sulfate, a t_p of 8°C (15°F) could not be maintained at high stoichiometry because of wet absorber discharge. As would also be expected from the high reactivity of this sorbent, the approach to saturation is not as critical to obtaining high levels of performance.

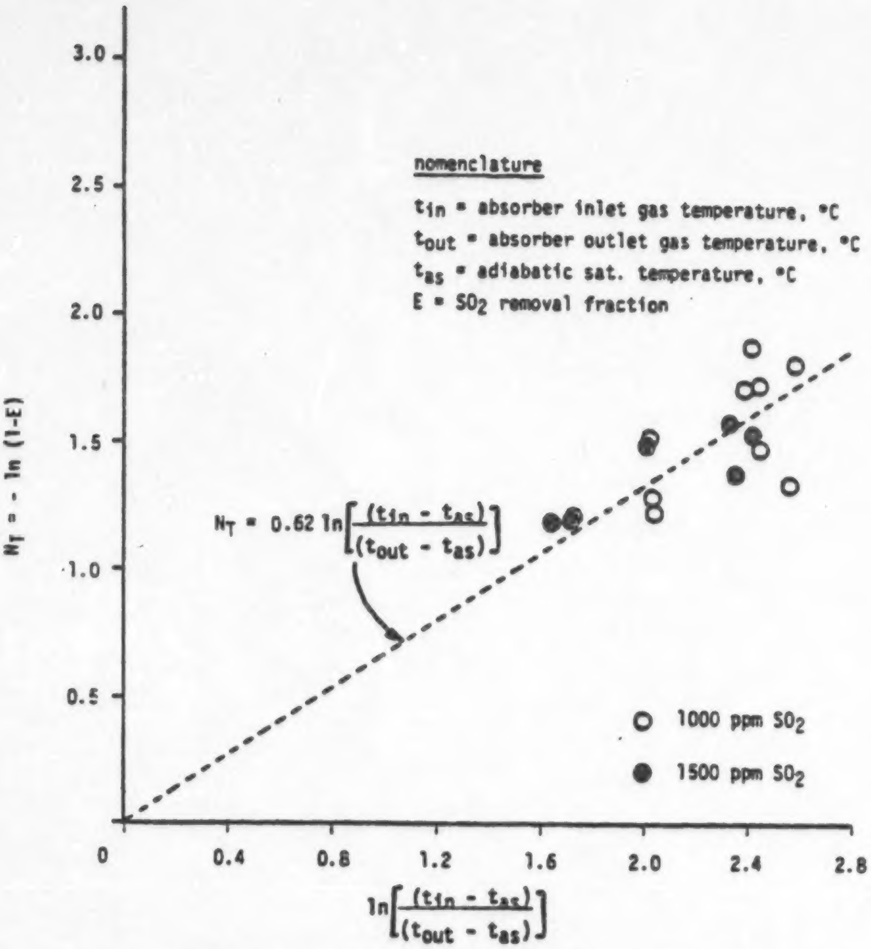


Figure 14A Comparison of measured SO_2 removal across absorber w/mathematical model for constant rate period. All data for straight-through lime, with stoichiometric ratio greater than 1.2 and residence time = 10 sec.

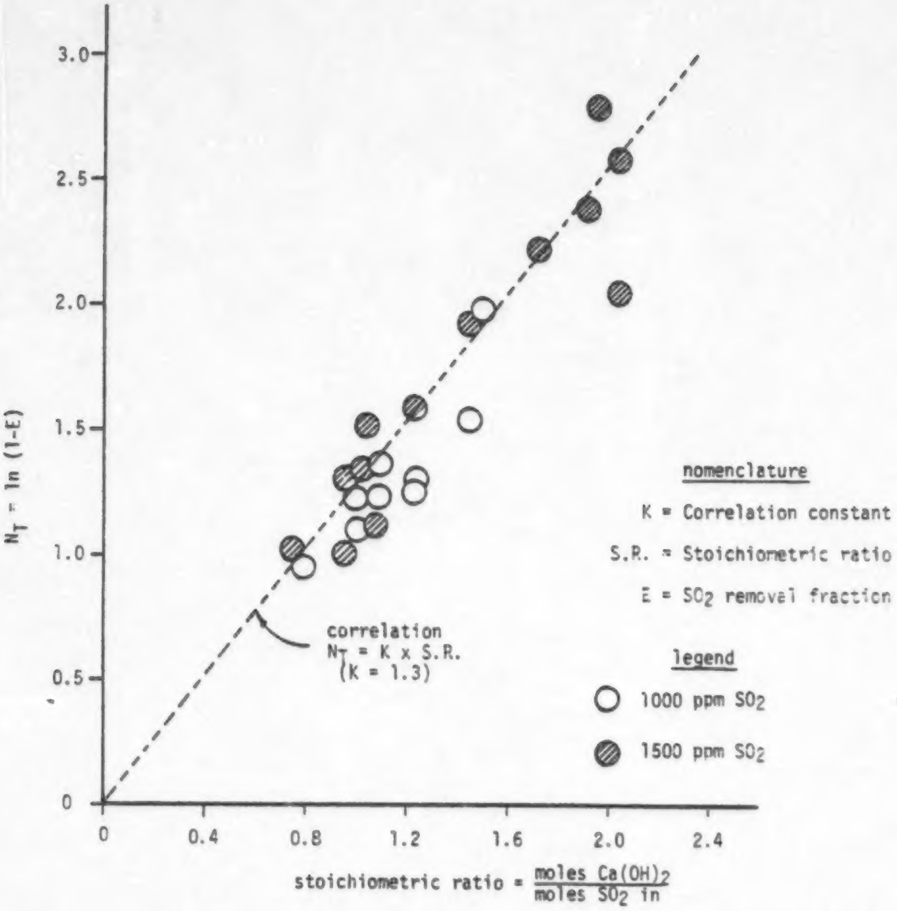


Figure 14B Application of correlation method based on mathematical model to removal data across system. All data for straight-through lime, 10° to 13°C (18° to 24°F) approach to saturation at the outlet, and residence time = 10 sec.

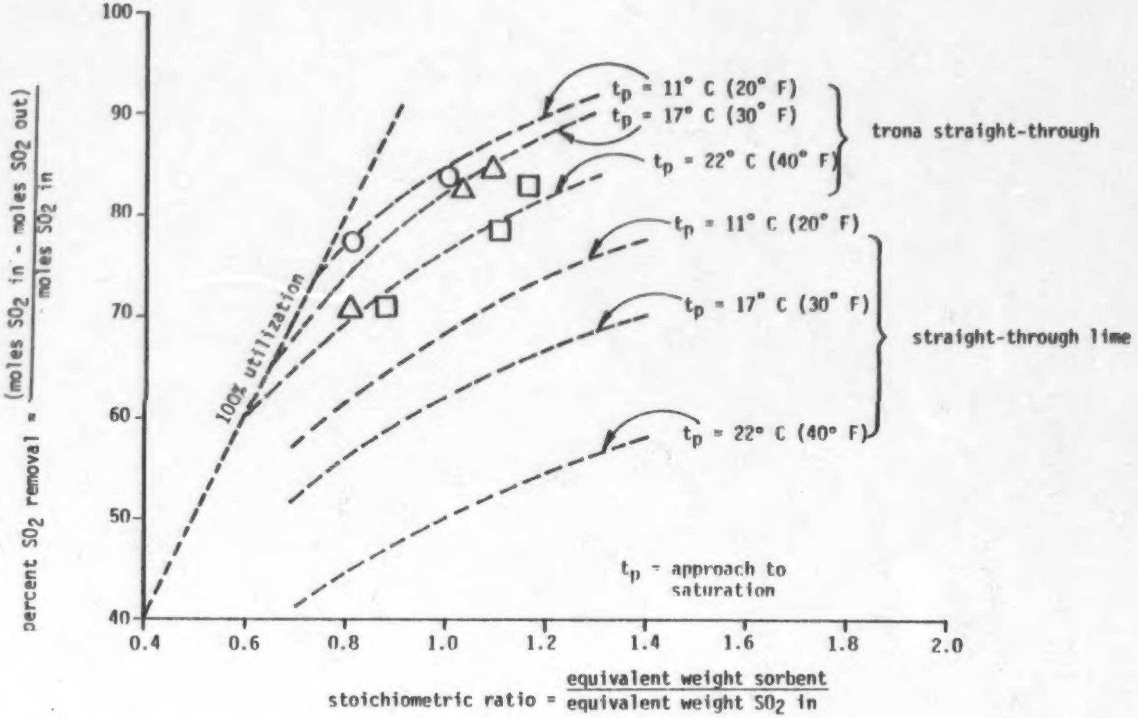


Figure 15 Comparison of SO_2 removal with trona and lime without off-product recycle for different approach temperatures, t_p, and residence time = 10 sec.

Limestone and Adipic Acid

A slurry of pulverized (Q4 grind) $CaCO_3$ was tested as a sorbent, with varying concentration of adipic acid additive. As may be readily seen from the data plotted in Figure 16, it is clear that adipic acid is effective in pepping up the relatively anemic reaction between the $CaCO_3$ and SO_2, but that performance is still far below that of lime, even at high t_p's.

Further tests at increased levels of adipic acid and with off-product recycle appear desirable in view of the low reactant utilization observed and the desirability of re-using residual adipic acid in the process off-product.

DRY INJECTION FGD TEST PROGRAM

Objectives and Scope

The dry injection tests were conducted to evaluate three low cost and reasonably available sodium compounds. SO_2 removal was measured over a representative range of gas temperatures, air/cloth ratios, and stoichiometric ratios, as summarized in Table 7. A more detailed description of all aspects of this test program is given in Reference 4.

Equipment Description

In addition to the fabric filter, the dry injection system consisted of a vibrating reactant storage hopper with a volumetric screw feeder, a rotary air lock, and a positive-pressure pneumatic conveying blower system (see Figure 17). Approximately 88 m^3/hr (52 acfm) of ambient air was used to convey the pulverized reactants to a counter-current injection nozzle located in a 45° elbow in the baghouse inlet duct.

Reactant Description

The reactants tested for FGD by dry injection were: nahcolite from the U.S. Bureau of Mines pilot mine in Horse Draw, Rio Blanco County, Colorado; and two grades of trona from Owens Lake in California (raw trona ore and trona upgraded to higher purity by mining methods). All dry injection reactants were tested in pulverized form, 95 percent minus 200 mesh. A summary of the chemical composition of the reactants used is given in Table 8.

Test Results and Discussion

The results of the dry injection tests, presented in Figures 18 through 20, indicate that nahcolite is by far the best performer, producing results comparable to lime spray absorption at 163°C (325°F) baghouse inlet temperature. Refined trona produced better results than the raw ore, indicating the strong effect of bicarbonate on SO_2 removal

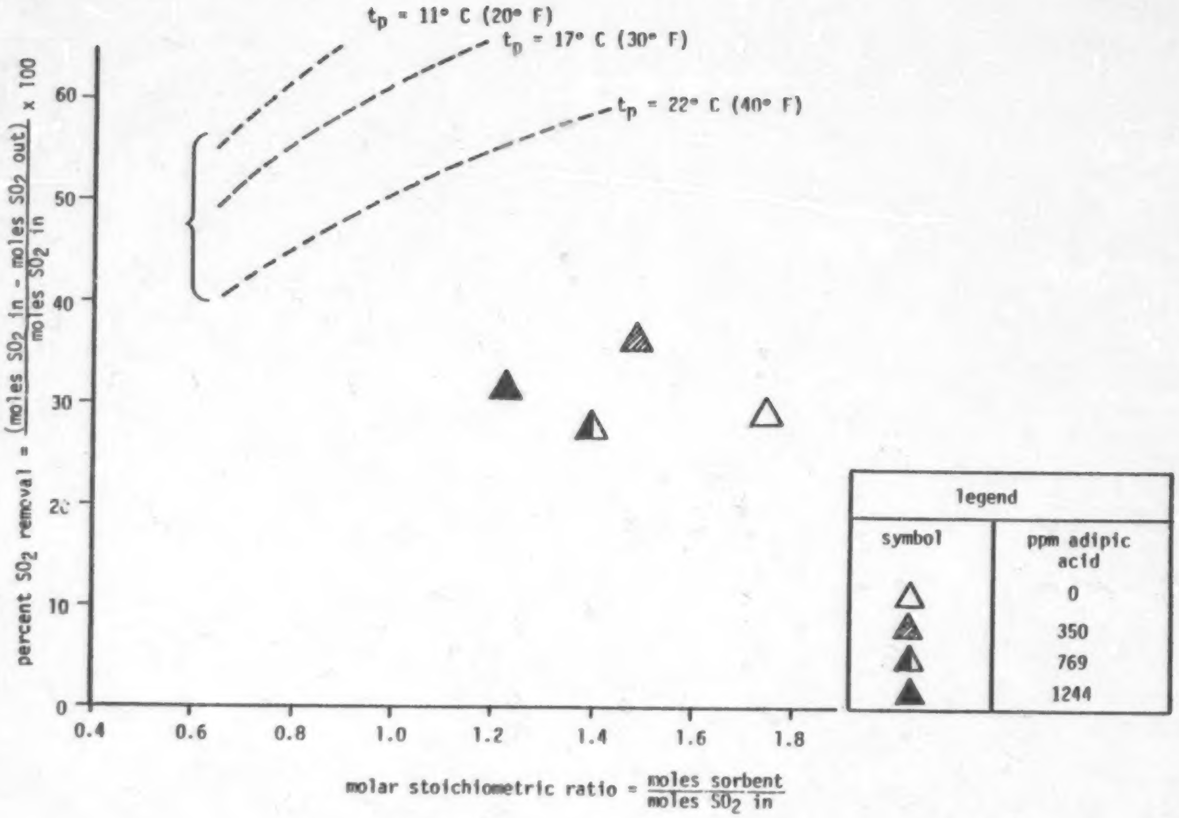


Figure 16 Comparison of SO_2 removal with lime and limestone plus adipic acid for no off-product recycle, 1000 ppm SO_2, approach temperature, t_p = 17°C (30°F) for limestone, and residence time = 10 sec.

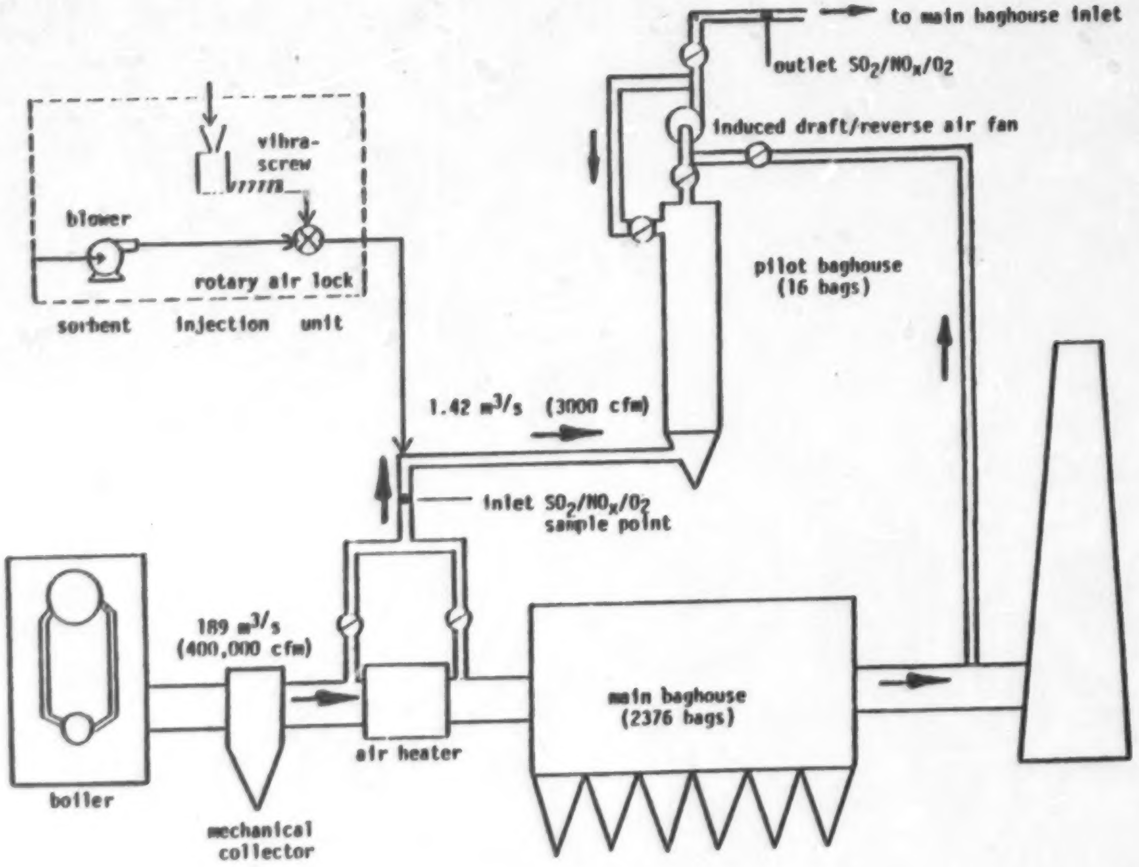

to main baghouse inlet
outlet $SO_2/NO_x/O_2$
induced draft/reverse air fan
vibra-screw
blower
rotary air lock
sorbent injection unit
pilot baghouse (16 bags)
1.42 m^3/s (3000 cfm)
inlet $SO_2/NO_x/O_2$ sample point
189 m^3/s (400,000 cfm)
main baghouse (2376 bags)
boiler
air heater
mechanical collector

Figure 17 Dry sodium injection/fabric filter SO_2 removal system on slipstream at City of Colorado Springs, Martin Drake Unit No. 6.

TABLE 7

Dry Injection Test
Variables and Their Ranges

Variable	Range
Sodium	Nahcolite, Trona, Refined Trona
Injection Temperature	204 to 327°C (400 to 620°F)
Baghouse Temperature	163 to 260°C (325 to 500°F)
Stoichiometric Ratio	0.7 to 2.1
Air-to-cloth Ratio	0.46 to 0.91 m/s (1.5 to 3.0 ft/min)

TABLE 8

Chemical Composition of Sodium Compounds Used in Dry Injection Tests

	Nahcolite		Refined Trona	Trona
Test Run	N-5A	N-5B	B-1	T-1
Constituent				
$NaHCO_3$	52.76	54.01	59.32	26.43
Na_2CO_3	3.64	3.13	14.22	41.39
NaCl	0.49	0.46	1.27	4.14
Na_2SO_4	0.40	0.32	1.19	5.19
Na_2SO_3	0.001	0.002	0.001	0.001
$NaNO_3$	40.86	40.01	15.34	9.72
Water	1.83	2.04	8.65	13.12

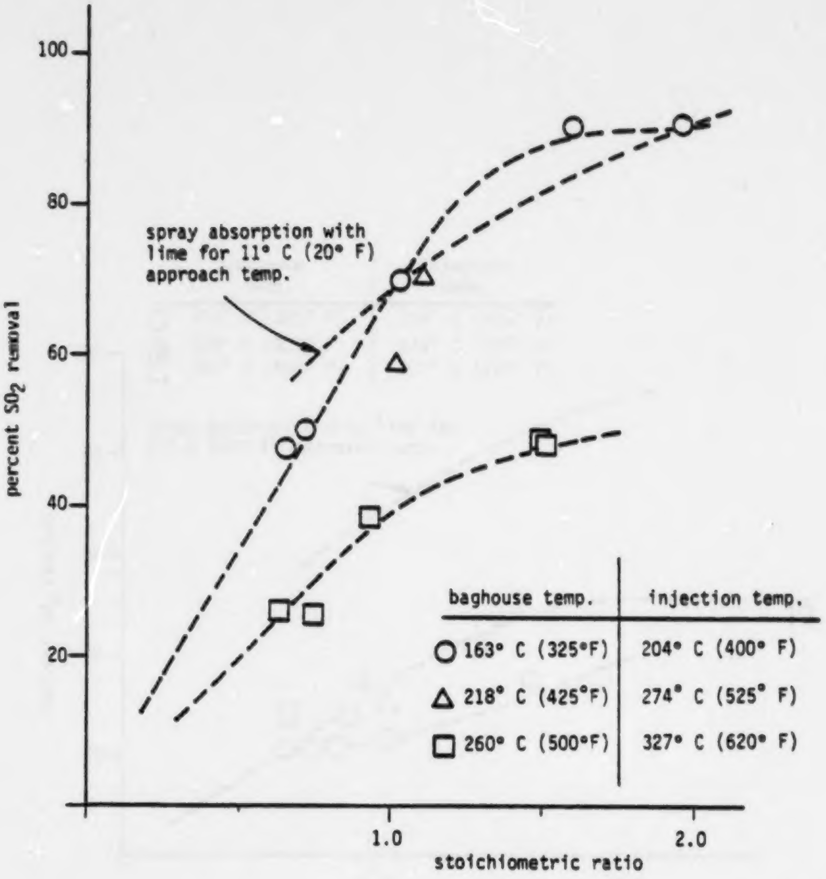


Figure 18 SO_2 removal with dry injection of nahcolite into duct of pilot system.

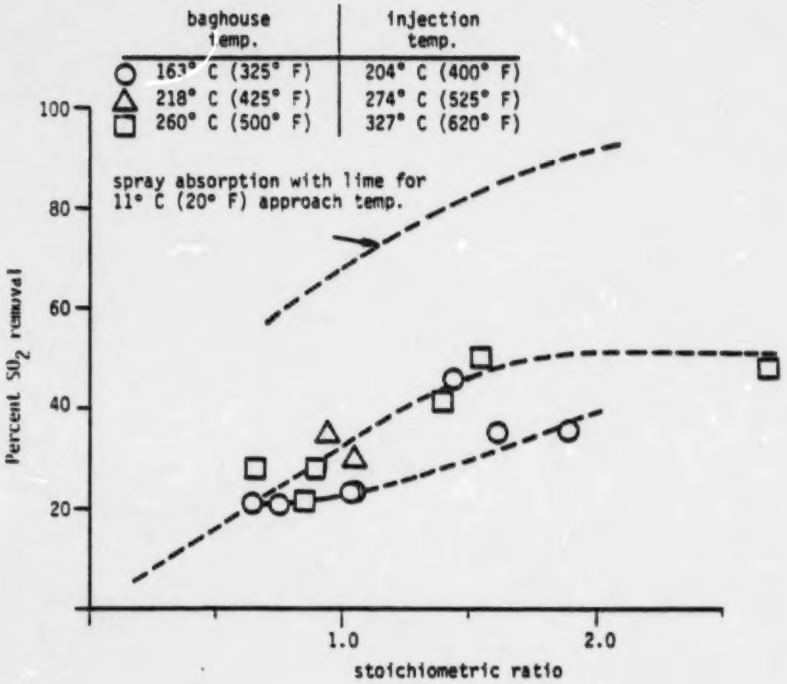


Figure 19 SO_2 removal with dry injection of trona into duct of pilot system.

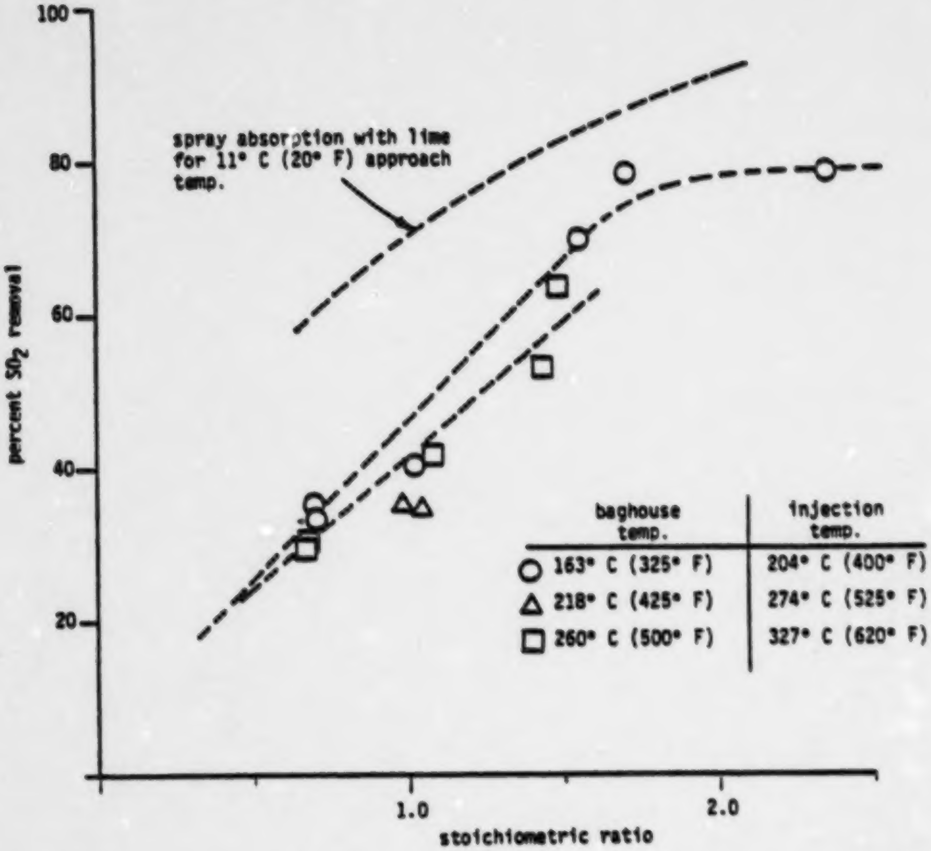


Figure 20 SO_2 removal with dry injection of refined trona into duct of pilot system.

performance. Gas temperature at the point of sorbent injection had a profound effect on removal, but air/cloth ratio was found to have little effect on performance.

Comparison with Spray Absorption Results

An estimated performance curve derived from the straight lime spray absorption performance data has been added to Figures 18 through 20 for comparison. It is apparent that only nahcolite offers performance comparable to lime-based spray absorption. None of the dry injection results, however, are even close to the levels of performance achieved with trona-based spray absorption.

SUMMARY

Spray absorption and dry injection as SO_2 control technologies were investigated in pilot-scale studies for EPA by Buell-Envirotech at the City of Colorado Springs' Martin Drake Station. In the spray absorption tests, the effect of inlet gas temperature and SO_2 concentration, reactant stoichiometry, and approach to saturation at the absorber outlet were varied. Each of these parameters affected SO_2 removal, but approach to saturation had the major effect as SO_2 removal was enhanced with decreasing approach. SO_2 removal with lime and recycled off-product reached 94 percent with a stoichiometric ratio of 2.0 and an approach temperature of 11°C (20°F). Recycling 50 percent of the off-product with lime increased SO_2 removal by up to 10 percent over that of lime alone for the Colorado coal fired. Decreasing the approach temperature from 17 to 8°C (30 to 15°F) increased SO_2 removal by about 15 percent for straight-through lime. As expected, SO_2 removal using trona exceeded that with lime, up to 50 percent more in tests without recycle of off-product for an approach temperature of 22°C (40°F) and a stoichiometric ratio of 1.2.

Three sodium compounds (nahcolite, trona, and refined trona) were evaluated as SO_2 absorbents in the dry injection studies, with gas temperature at injection location, fabric filter compartment (baghouse) temperature, stoichiometric ratio, and air-to-cloth ratio as parameters. All three compounds were effective at removing SO_2 from the slipstream flue gas. Nahcolite was the best performer, yielding 70 percent SO_2 removal at a stoichiometric ratio (S.R.) of 1.05 and 91 percent removal at S.R. = 2.0. Refined trona outperformed trona, but its SO_2 removal capability was considerably below that of nahcolite. While SO_2 removal with nahcolite and refined trona decreased as gas temperature at the point of injection increased from 204 to 327°C (400 to 620°F), the opposite was true with trona. Air-to-cloth ratio over the range from 1.44 to 2.99 had no apparent effect on SO_2 removal.

ACKNOWLEDGEMENTS

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LIST OF ABBREVIATIONS

acfm	--	Unit of flow volume, actual cubic feet per minute
atm	--	Unit of pressure, atmosphere
°C	--	Unit of temperature, degrees Celsius
°F	--	Unit of temperature, degrees Fahrenheit
ft/sec	--	Unit of velocity, feet per second
FGD	--	Flue gas desulfurization
g	--	Unit of mass, gram
kcal	--	Unit of heat, kilocalorie
kg	--	Unit of mass, kilogram
l	--	Unit of volume, liter
m	--	Unit of length, meter
mg	--	Unit of mass, milligram
ml	--	Unit of volume, milliliter
mm	--	Unit of length, millimeter
mm Hg	--	Unit of pressure, millimeters mercury column
m^3/hr	--	Unit of flow volume, actual cubic meters per hour
m/sec	--	Unit of velocity, meters per second
pH	--	$-\log_{10}[H^+]$, H^+ = hydrogen ion concentration, g-moles per liter
ppm	--	Unit of concentration, parts per million by volume for gases, by mass for liquids and solids
rpm	--	Unit of angular velocity, revolutions per minute

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Dry scrubber demonstration plant — operating results

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Application and acceptance of dry scrubbing to remove SO_2 from boiler flue gas has continued at a rapid pace since the first small pilot plants were built several years ago to investigate the concept. And now, the results from the first large demonstration plant are available.

B&W's first dry scrubbing pilot project was an 8000 acfm unit constructed in 1978 at Basin Electric's Neal Station in Velva, North Dakota. Tests were also conducted at B&W's Alliance Research Center with a 1500 acfm pilot built in 1979. Both of these efforts have been previously reported.[1,2] In parallel with these activities, B&W laid plans in 1978 to construct a 20 MW dry sulfur removal (DSR) demonstration unit. Pacific Power & Light Company agreed to provide a site at their Jim Bridger Station Unit 3 which burns a low-sulfur, Western bituminous coal. Operation and testing of this demonstration unit are nearing completion.

Description of demonstration facility

Construction of the DSR demonstration facility began in spring, 1979. The basic components of the system, shown schematically in Figure 1, are a reactor with ID booster fan and flues, a lime slaking building with ash recycle equipment, a pilot precipitator, and a pilot baghouse. Specifications for the demonstration facility are given in Table 1. A view of the overall system is shown in Figure 2.

Flue gas for the demonstration facility was taken from the north air preheater outlet on the Unit 3 boiler. Four large scoops were placed across the preheater discharge fluework to isokinetically sample flue gas and dust for the demonstration unit.

The reactor concept depicted in Figure 3 was based on the pilot plant designs previously used at the Neal Station and the Alliance Research Center. Featured in this design is a horizontal flow reactor utilizing dual fluid atomization, analogous to the combustion of liquid fuels. In both processes, major emphasis is placed on fine atomization and thorough mixing of the atomized particles with flue gas. A photograph of the atomizer assembly taken from inside the reactor is shown in Figure 4. Flue gas from the reactor is ducted back to Unit 3 at the main unit precipitator inlet.

The pilot precipitator and baghouse are designed to operate on a 3000 acfm slipstream taken from the discharge flue of the reactor. Although the baghouse and precipitator have many features of commercial equipment, their purpose was to demonstrate process performance rather than the operability of commercially available equipment. The baghouse utilizes full-sized, Teflon-coated glass fiber filter bags. Bags are cleaned by reverse flow deflation. The precipitator is a Rothemuhle three-field, rigid-frame electrode design.

Lime slaking is accomplished in a 4 ft x 8 ft wet ball mill. Product from the ball mill is diluted with

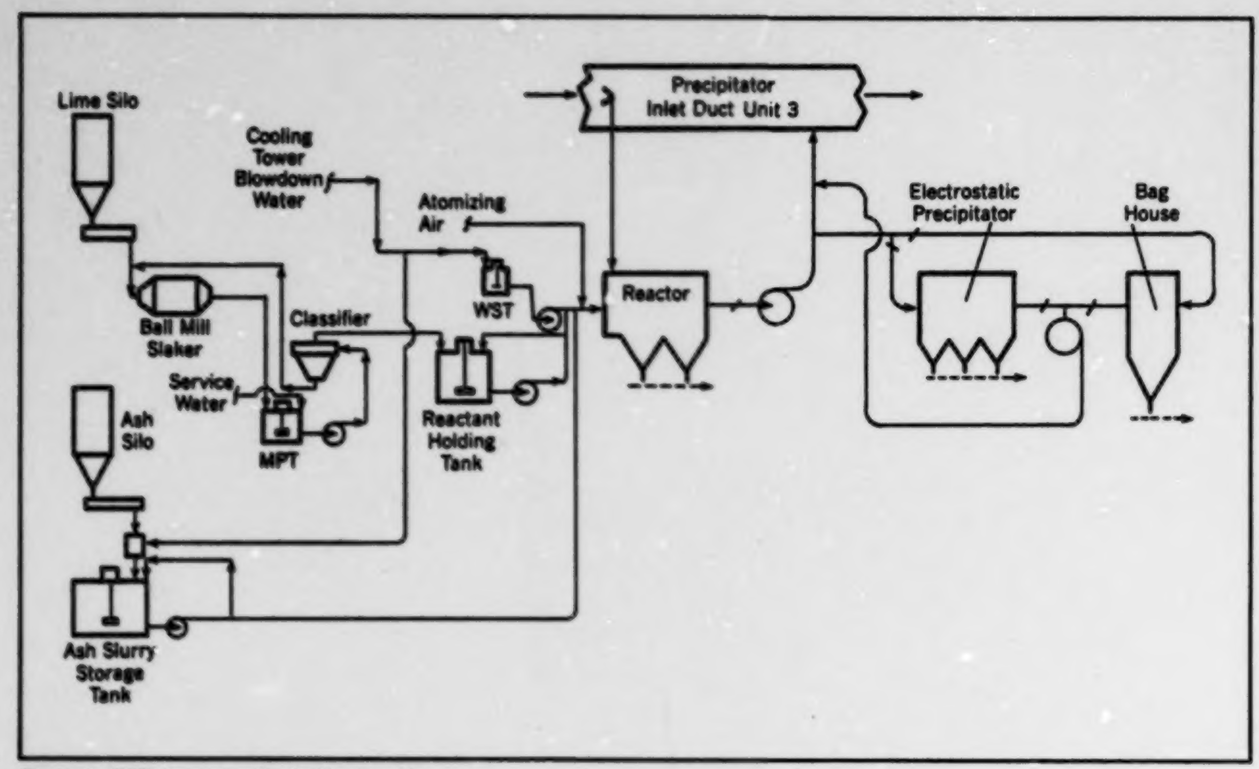

Figure 1 DSR demonstration facility flow diagram

low-solids plant water and classified in a hydroclone with the underflow returning to the inlet of the mill and the product overflow going to the slurry feed tank. Equipment also is available to slurry and feed fly ash, partially spent recycled material, soda ash, and soda liquor.

Control systems for the reactor, baghouse, and precipitator are virtually identical to those that would be employed on commercial systems. For the reactor, the two major control loops are for the slurry and dilution water flow.

The amount of slurry feed required is determined from a stoichiometric calculation. The rate of SO_2 is determined from the product of the flue gas flow multiplied by the inlet SO_2 concentration; and the rate of lime is determined from the slurry feed flow multiplied by the slurry density. The controls then regulate the feed slurry valve to produce the desired lime feed rate for a specific operating condition.

Dilution water flow to the reactor is regulated to produce a set point approach to saturation temperature for the flue gas leaving the reactor. This regulation is accomplished by metering the gas flow and water flow to the reactor and calculating the quantity of water required to produce the desired temperature. A control signal modulates the dilution water valve to produce the desired temperature. This temperature is measured at the reactor outlet to produce a trimming signal for final control adjustment.

SO_2 is monitored with a Dupont Model 411 Ultraviolet Analyzer. Extractive samples are taken from the fluework at the reactor inlet and outlet and from the precipitator and baghouse outlet. Gas samples are drawn through heated, insulated tubing to a gas conditioning unit for final filtration and water knockout before passing through the SO_2 analyzer.

A parallel sample is taken from the gas conditioner to a Beckman oxygen analyzer. Sample connections are located in fluework adjacent to the gas sample probes so that dust loadings can be determined. In addition, a Lear Siegler opacity meter is located in the discharge fluework from the precipitator and baghouse.

A small laboratory facility is located on site to make routine analyses of the lime, lime slurry, and spent reactant materials. Data logging and reduction are done on the main computers located in Barberton, Ohio, and the Alliance Research Center.

Operations

Calibration and shakedown of the demonstration unit began in August, 1979, and after a six week

Table 1
DSR demonstration unit specifications

Reactor Specifications	
Type	Horizontal
No. of atomizers	6 and 1
Atomizer type	Y-jet air atomizing
Flow volume	120,000 acfm
Draft loss	2.5 in WG
Reactor retention time	6 sec
Reactor flue gas velocity	5 ft/sec
Number of hoppers	2
Length	40 ft 5 in.
Width	18 ft
Precipitator Specifications	
Number of fields	3
Field height	7.2 ft
Field depth	6.5 ft
Lane spacing	12 in.
Collecting surface/field	283 ft^2
Total collecting surface	848 ft^2
Electrode length/field	157 ft
Total electrode length	471 ft
Aspect ratio	2.7
Cross sectional area	21.3 ft^2
Gas volume	3000 acfm
Gas velocity	2.4 ft/sec
Specific collection area	282 ft^2/1000 cfm
Filterhouse Specifications	
Physical arrangement	
Number of compartments	3
Bags per compartment	5
Bag details	
Diameter	11.5 in.
Length	30-36 ft
Cleaning	Reverse gas flow
Controls	Programmable logic controller
Capacity	
Gas volume	
(at 2:1 a/c filtering ratio and with all compartments in operation)	2670 acfm 30 ft bags 3210 acfm 36 ft bags
Ball Mill Specifications	
Length	8 ft
Diameter	4 ft
Power	40 hp
Ball charge	8200 lb
Single classifier	95 through 325 mesh
Closed loop system	

boiler outage, lime slurry testing was begun in November of that year. The objectives for operation and testing of the DSR demonstration unit were to: 1) demonstrate on a large scale the operation of a multiple atomizer reactor utilizing concepts developed in liquid fuel combustion processes, 2) demonstrate the controllability of the process in utility boiler application, 3) demonstrate equipment reliability and performance, 4) confirm the performance parameters measured previously in pilot testing, and 5) demonstrate the operation of a single, commercial-size atomizer. After approximately 1750 hours of operation and testing, these objectives have been clearly attained.

Modifications have been required to achieve these results. Significant among these modifications are the following:

1. In the initial configuration, plant steam from the Unit 3 boiler was used for slurry atomization. Although steam produced good atomization, it also led to a rapid buildup of calcium deposits inside the atomizer sprayer plates at the point where slurry and steam first contact each other. These deposits occurred very rapidly, were extremely hard, and were layered. It is believed that soluble calcium was rapidly precipitated when 140 F slurry contacted the atomizing steam at 400 F. Because of no ready solution to this problem, the atomizing medium was converted from steam to compressed air.
2. As originally designed, the gas side pressure loss across the atomizer throats was 1.5 in. H_2O. In order to duplicate pilot plant performance, it became necessary to increase gas side mixing at the atomizer throat. To accomplish this, additional throat vanes were installed with an increase in gas side resistance to a total of 2.5 in. H_2O.
3. To keep any mist impingement on reactor surfaces under all operating conditions, the average residence time was increased from 6 to 9 seconds.

The ball mill slaking system has largely been automated, and although it operates smoothly, operator attendance for minor adjustments is required. Solids leaving the mill are controlled at 38 to 40 percent by weight. This concentration produces a fine, reactive material with minimum water requirements.

To achieve 40 percent solids leaving the mill, makeup water is added only at the mill discharge. Slaking water is provided at the front of the mill from the classifying hydroclone underflow which is recycled from the mill outlet. The hydroclone overflow at 25 percent solids goes directly to the slurry feed tank. Since tests indicated that use of high-solids cooling tower blowdown water for slaking increased the slurry particle size and was detrimental to reactor performance, low-solids plant water is used for slaking.

The slaking system is giving good, consistent results as measured by uniformly low settling rates of the final product and little or no wear on pumps, control valve parts, or atomizer sprayer plates.

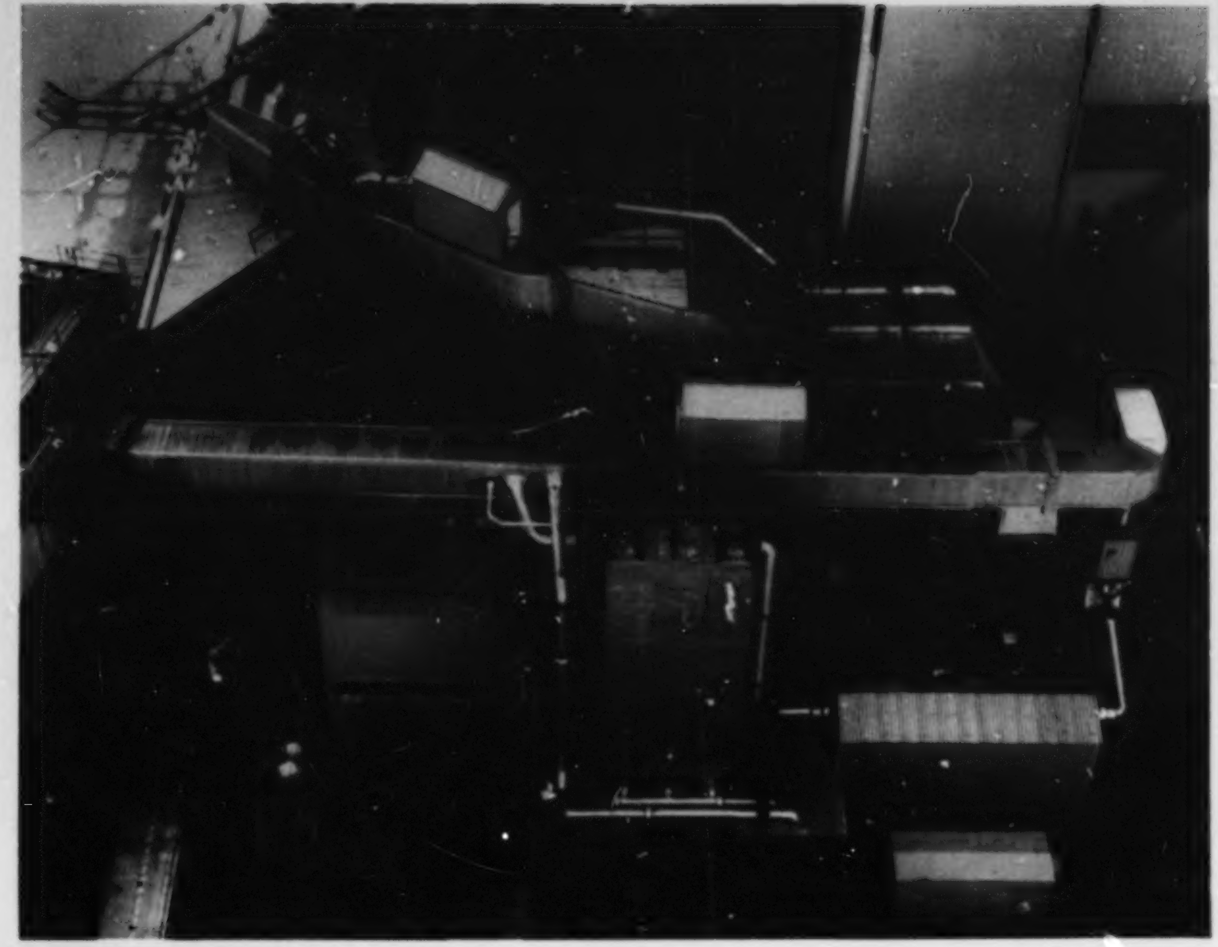

Figure 2 Overall view of demonstration facility

Although the hydroclone's product cut size is 95 percent minus 325 mesh screen, an automatic backflush filter is located in the feed slurry line to catch any debris or large unclassified particles.

Dilution water for temperature control is stored in a separate feed tank and pumped separately for mixing with the lime slurry at the atomizer front. Internal feed line deposits have been effectively controlled by raising the pH in the dilution water feed tank to 10.5 with the addition of small amounts of lime.

The automatic outlet temperature control system, which modulates the dilution water control valve, has done an unusually good job of controlling the reactor outlet temperature at a set point approach to adiabatic saturation temperature. Load changes of 25 percent per minute have been simulated with only a 2 degree temperature swing at the reactor outlet. Reactor inlet temperatures also have been

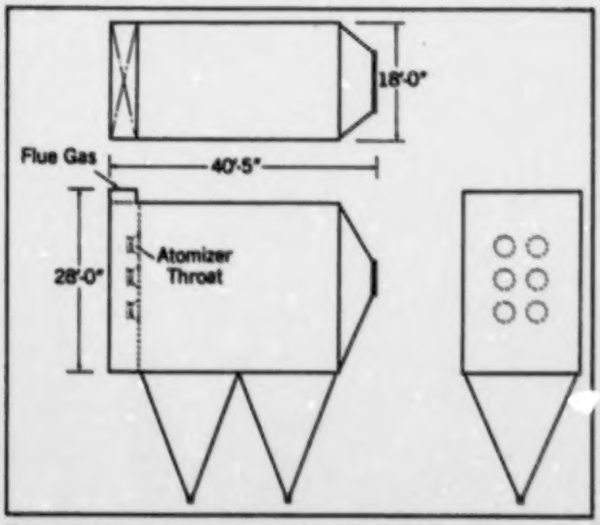

Figure 3 Horizontal reactor for demonstration facility

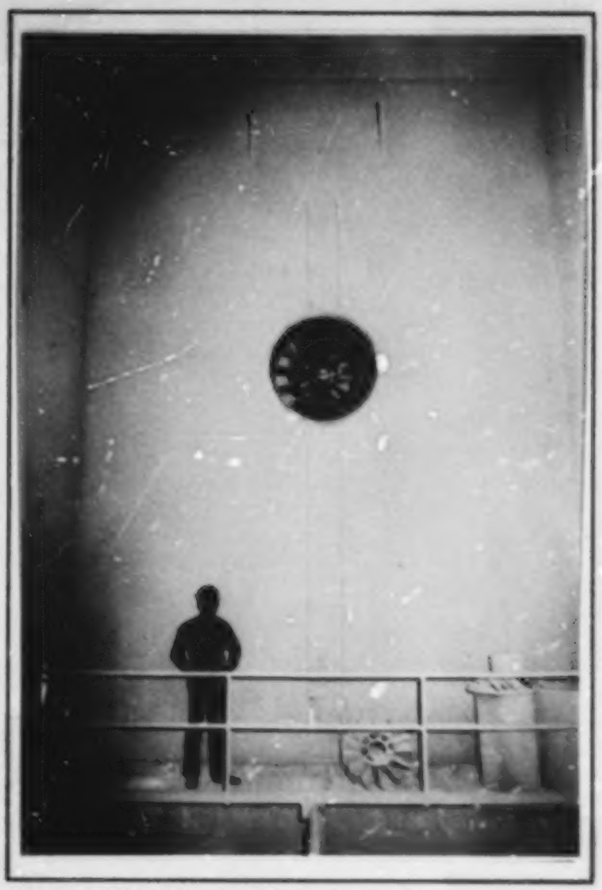

Figure 4 Inside DSR reactor facing front wall

lowered at 25 degrees per minute with the same results.

Atomizer sprayer plate wear, when spraying lime slurry, has been virtually undetectable with sprayer plates made of titanium carbide or tungsten carbide materials. When ash slurries are used, however, fairly rapid wear has been observed with those materials.

To rapidly test sprayer plates made of other hardened materials, a special wear loop was constructed. Of the several materials tested, silicon carbide and silicon nitride have shown suitable life. Figure 5 shows a silicon carbide tip after 300 hours of continuous operation with no measurable wear. This material is expected to give a minimum of six months operating life with an abrasive ash slurry.

System performance

Tests at the Neal Station and B&W's Alliance Research Center have identified that stoichiometric ratio, alkalinity of the coal ash, temperature approach to adiabatic saturation (T_{as}), as well as inlet gas temperature to the reactor (T_{in}) have an important effect on SO_2 removal efficiency. The primary variable for data correlation, however, continues to be stoichiometric ratio defined as the pound moles of alkali, such as lime, per pound mole of entering sulfur dioxide.

At an inlet gas temperature of 280 F and an outlet temperature 20 F above adiabatic saturation temperature, the reactor performance with Jim Bridger coal is shown in Figure 6. During these tests, the SO_2 inlet concentration was in the range of 350 to 440 ppmvd. When spraying with water alone, the intercept is shown at about 13 percent SO_2 removal. This value corresponds closely to previous work at our research center pilot when burning the same Jim Bridger coal. Typical analysis of as-fired coal and of ash taken from the reactor inlet are shown in Table 2.

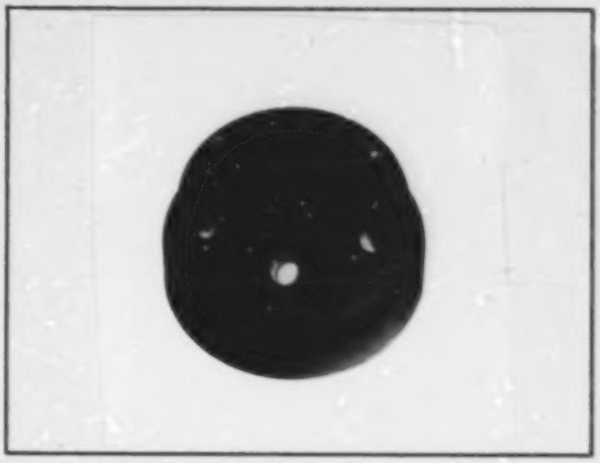

Figure 5 Silicon carbide atomizer tip

Table 2
Jim Bridger Station
Wyoming bituminous coal
(Jim Bridger mine)

Coal analysis (%)		Ash analysis (%)	
Carbon	68.00	Calcium as CaO	6.10
Hydrogen	4.50	Magnesium as MgO	2.60
Nitrogen	1.29	Sodium as Na_2O	3.49
Sulfur	0.70	Potassium as K_2O	1.05
Oxygen	12.81	Sulfur as SO_3	0.69
Ash	12.70	Silicon as SiO_2	60.08
	100.00	Aluminum as Al_2O_3	17.81
Total moisture (%)	19.7	Iron as Fe_2O_3	3.03
Btu per lb (wet)	9480	Titanium as TiO_2	1.16
		Carbon	2.90

Additional tests were run under the same operating conditions with the precipitator and baghouse in operation. The overall efficiencies from

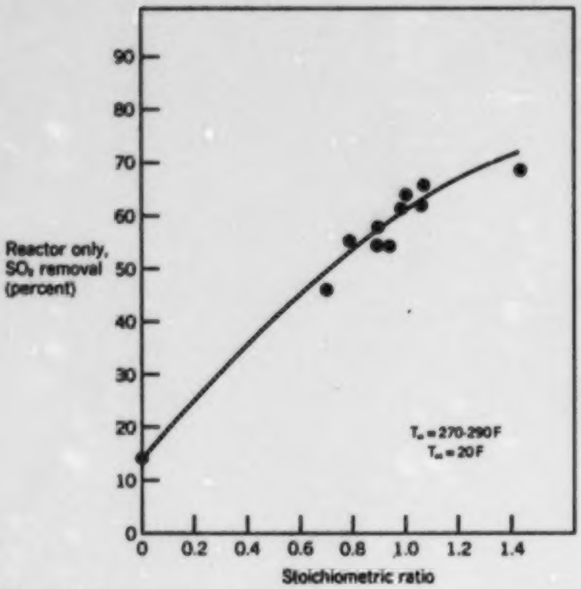

Figure 6 Reactor only performance

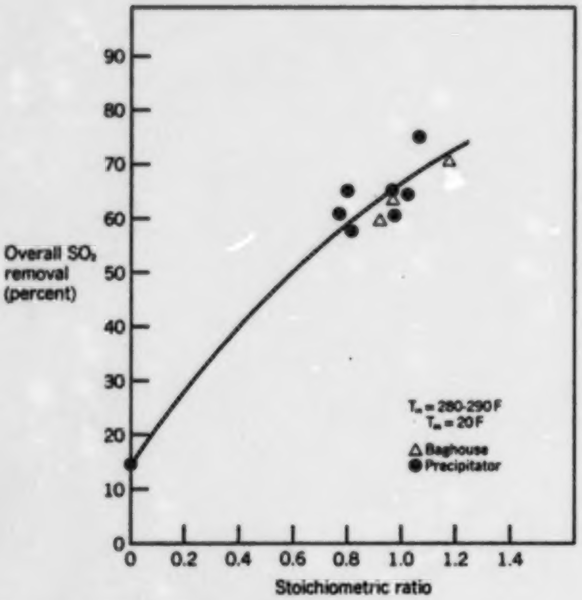

Figure 7 Overall system performance

these tests, plotted in Figure 7, show that both particulate collecting devices gave about the same overall performance and removed 15 to 25 percent of the SO_2 entering the collection device.

Previous pilot tests conducted by B&W have shown that the alkaline material in fly ash can be utilized effectively in the dry scrubbing process. To demonstrate this effect on a large scale, highly alkaline fly ash from Basin Electric Company's Laramie River Unit 1 precipitator was shipped to the Jim Bridger Station for tests in the DSR unit. Coal for the Laramie River Station is supplied by the Cordero mine in the Powder River Basin. Analysis of the Laramie River ash is shown in Table 3. Provisions were made to pneumatically inject fly ash into the gas stream ahead of the DSR reactor and to slurry the fly ash for direct atomization into the reactor. Results from these tests are plotted in Figure 8. In this case, the intercept when spraying a slurry of Laramie River ash alone shows about 65 percent SO_2 removal. These tests indicate that the dry scrubbing process can utilize high alkalinity ashes, and for some sites it can meet the compliance requirements with a minimum purchase of lime.

Test data indicates that recycling partially spent reactant and ash can likewise significantly benefit the lime utilization. In Table 4, alkalinity analyses are shown for the material caught in the reactor hoppers and the three baghouse hoppers without recycle. As seen from this data, the system has classified the particulate with the high alkalinity fraction collected in the particulate device and the low alkalinity fraction in the reactor hoppers. The ratio of total alkalinity in this case is approximately 3:1 and demonstrates that reinjection of material from the particulate collection device rather than the reactor can be beneficial. Material collected from the particulate collector hoppers was slurried and sprayed into the reactor along with lime slurry. Under comparable conditions, the SO_2 removal efficiency with recycle increased from 70 to 86 percent.

Table 3
Laramie River Station
Wyoming subbituminous coal
(Cordero mine, Powder River Basin)

Ash analysis (%)	
Calcium as CaO	20.27
Magnesium as MgO	3.45
Sodium as Na_2O	1.18
Potassium as K_2O	0.31
Sulfur as SO_3	11.71
Silicon as SiO_2	35.58
Aluminum as Al_2O_3	16.82
Iron as Fe_2O_3	4.73
Phosphorous as P_2O_5	0.89

As expected, the inlet gas temperature to the reactor and the outlet temperature from the reactor, expressed as the temperature difference between the outlet temperature and the adiabatic saturation temperature, have a definite effect upon SO_2 removal efficiency. As shown in Table 5, a 10 degree increase in the temperature leaving the reactor (from a 20 degree approach to saturation to a 30 degree approach to saturation) decreased the

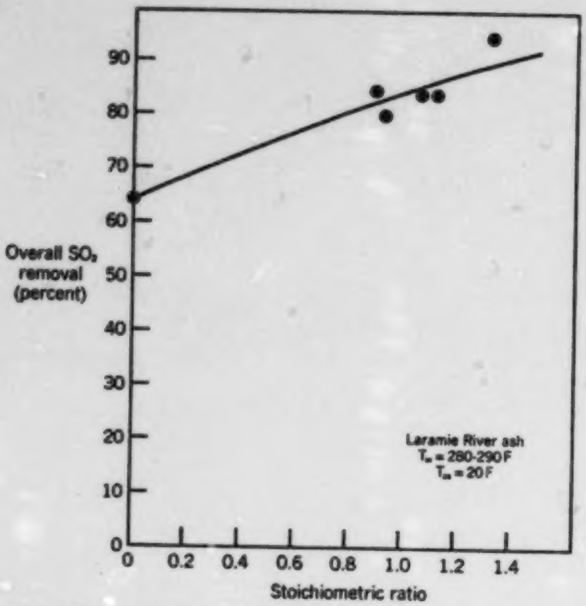

Figure 8 High alkalinity fly ash performance

Table 4
Hopper ash analyses comparison

	Reactor hoppers		**Particulate collector hoppers**		
	A	B	A	B	C
Calcium (Ca)	5.50	4.60	14.90	17.90	19.30
Magnesium (Mg)	0.01	0.56	0.02	0.98	0.80
Potassium (K)	1.50	1.50	1.50	1.50	1.50
Sodium (Na)	0.50	0.50	0.70	0.70	0.70

Table 5
Data showing temperature effects

Reactor inlet gas temperature	**Reactor outlet gas temperature**	**Approach to saturation**	**Overall removal efficiency**
280 F	138 F	20 F	75%
280 F	148 F	30 F	65%
230 F	138 F	20 F	65%

SO_2 removal efficiency by 10 percentage points compared to the base condition. Likewise, a reduction of the inlet gas temperature by approximately 50 degrees resulted in a decrease in SO_2 removal efficiency of 10 percentage points. These operating characteristics have been observed in the two B&W pilot units and the relationships are under continuing study.

Tests were conducted to establish the effect of atomizer removal while the remainder of the system continued in operation. One of the six atomizers was removed from operation by shutting off the slurry and atomizing air flow. The gas flow, however, was not shut off. Thus, untreated flue gas entered the reactor through the throat whose atomizer was out of service.

These tests were repeated with each of the other atomizers shut off, and in all cases the results indicated no effect on the SO_2 removal efficiency. Additional tests were run to see the effect of atomizer spacing or spray overlap. Alternate atomizer throats were blanked off and the atomizers were removed from service. During these tests, no change in the reactor SO_2 removal efficiency was measured.

Results from tests with commercial soda ash gave higher overall utilizations compared with the lime tests and show that soda ash would be a desirable reagent if not for the material's high cost. The results of the soda ash tests, plotted in Figure 9, were conducted under less than favorable conditions in that the inlet gas temperature ranged between 235 and 250 F as a result of low boiler load during the test period. As with lime, the approach to saturation temperature at the outlet of the reactor had a similar influence on SO_2 absorption.

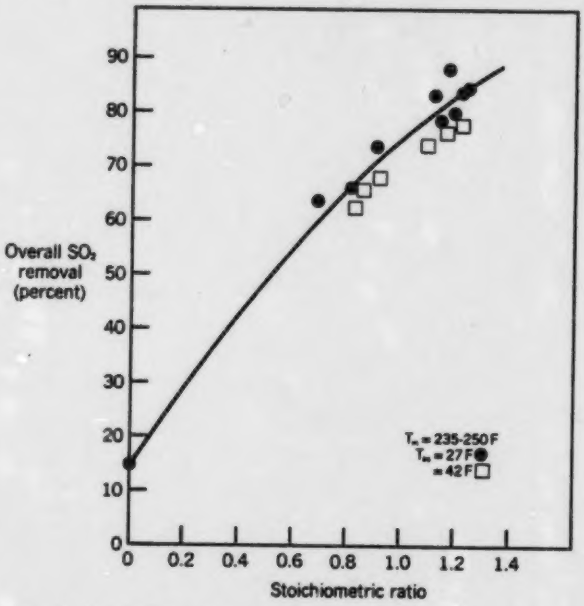

Figure 9 Soda ash performance

Summary

Large-scale demonstration plant operation and testing under actual operating conditions have proven to be an essential step toward commercialization of the DSR process. Integrating the various components of the process into a completely automated system could not have realistically been accomplished in small pilot scale.

Knowledge gained from the Jim Bridger DSR demonstration unit has been applied to the 570 MW DSR unit under construction at Basin Electric's Laramie River Unit 3 scheduled for operation in fall, 1981, and Colorado Ute Electric Association's Craig Unit 3 scheduled for operation in fall, 1982. Both of these dry scrubbers are to be installed in plants where wet scrubbers were installed on earlier units for SO_2 control. In these cases, the dry scrubber was perceived to be a better economic and operational choice to accomplish compliance to SO_2 emission standards.

References

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2. Downs, W., Sanders, W. J., and Miller, C. E., "Control of SO_2 emissions by dry scrubbing," American Power Conference, Chicago, Illinois, April 21-23, 1980.

Session 7: INDUSTRIAL APPLICATIONS

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Applicability of FGD Systems
to
Industrial Boilers

by

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ABSTRACT

The Clean Air Act Amendments of 1977 require the Environmental Protection Agency (EPA) to coordinate and lead the development and implementation of regulations on air pollution. These include standards of performance for new and modified sources of pollution. Specifically mentioned as a prioritized pollution source in the August 21, 1979 Federal Register are industrial fossil fuel-fired steam generators. Accordingly, the EPA has undertaken a series of studies of industrial boilers and pollution control systems with the intent to promulgate standards of performance based on the study results.

This paper presents the results of an evaluation of the application of flue gas desulfurization (FGD) controls for industrial boilers. Factors considered included development status, environmental impacts, energy impacts, and capital and operating costs of the various FGD processes. The focus of this paper will be to present the results of the environmental, energy, and cost impact analysis, and will consider only those processes that have been commercially applied to date.

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APPLICABILITY OF FGD SYSTEMS TO INDUSTRIAL BOILERS

INTRODUCTION

The Clean Air Act Amendments of 1977 require the Environmental Protection Agency (EPA) to coordinate and lead the development and implementation of regulations on air pollution. These include standards of performance for new and modified sources of pollution. Specifically mentioned as a prioritized pollution source in the August 21, 1979 Federal Register are industrial fossil-fuel-fired steam generators. Accordingly, the EPA has undertaken a series of studies of industrial boilers and pollution control systems with the intent to promulgate standards of performance based on the study results.

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There are currently more than 130 operating FGD units being used to control SO_2 emissions from industrial boilers.[1] These applications range in size from units treating gases from boilers of about 25 to 800×10^6 Btu/hr. The majority of industrial boiler FGD systems currently in use are of the once-through sodium scrubbing design (102 operating units). This design apparently has been favored due to its low capital cost, overall simplicity and high reliability. Other FGD systems currently being used for industrial boiler applications include dual alkali, spray drying, and lime and limestone systems. Table 1 illustrates the distribution of industrial boiler FGD systems currently in use in the United States.

TABLE 1. DISTRIBUTION OF OPERATING INDUSTRIAL BOILER FGD UNITS

Process Type	No. of Units	Percent of Total
Sodium Scrubbing	102	76
Dual Alkali	21	16
Lime and Limestone	2	1.5
Spray Drying	2	1.5
Others	7	5

ENVIRONMENTAL IMPACTS

The air, liquid, and solid waste impacts of the various FGD systems applied to industrial boilers were considered as functions of system size, SO_2 removal level, and fuel sulfur content. With regard to air pollution, each of the FGD systems has the ability to remove both particulates and SO_2, however, only SO_2 removal has been considered. The impact of all the systems with respect to SO_2 emissions is the same since each of the processes can be designed to achieve the same degree of SO_2 control. A possible exception to this is the lime spray drying process which may not be able to achieve 90 percent SO_2 removal on high sulfur coals. Although supporting data have not been received, some process vendors claim that 90 percent SO_2 removal on high sulfur coals is achievable with their spray drying systems.

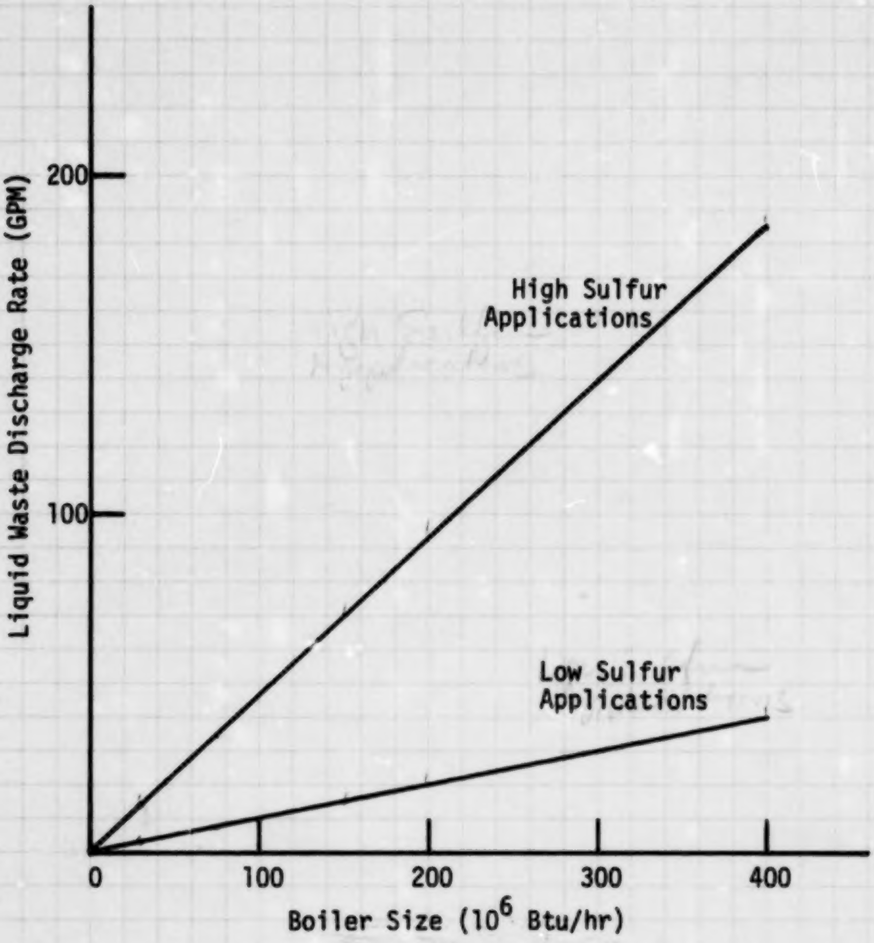
Liquid Waste Impacts

With regard to water pollution, only the once-through sodium (NA) systems result in direct liquid discharges. Aqueous wastes from these processes contain dissolved sodium sulfite/sulfate salts. The spray drying process is designed to produce a dry waste product and, thus, will have no liquid waste stream for disposal. The dual alkali (DA) and lime and limestone systems can be designed to operate closed-loop so that the only water losses during normal operation occur with the sludge going to landfill and by evaporation out the stack. Any purging of these systems due to water imbalances or other operational upsets, system blowdown to prevent scaling, or operator error will result in the discharge of an aqueous waste stream which can be contained and treated. However, during normal operation, there should be no water pollution impact from spray drying, lime and limestone, and DA FGD systems designed on a closed-loop basis.

Figure 1 illustrates the amount of aqueous wastes generated from sodium scrubbing systems as a function of size for both high sulfur and low sulfur coal applications. Discharge rates range from about 10 to 200 gpm for high sulfur coal applications and from about 3 to 40 gpm for low sulfur applications.

Dissolved solids and pH-imbalances are the two main areas of concerns for which treatment may be required for wastes from a sodium scrubbing system. Discharge to an evaporation pond or to an existing centralized wastewater treatment facility is commonly practiced. Of 102 sodium scrubbing systems in use today, about 80 use evaporation ponds (over 30 of these in conjunction with well injection), and 10 use centralized water treatment for disposal of FGD wastes.[2]

If the scrubber effluents are being discharged directly to a receiving stream, the water quality standards applicable to that stream will govern the degree of treatment required. Also, if the scrubber effluents are discharged to a publicly owned treatment works (POTW), then the pretreatment requirements contained in the guidelines for that POTW will determine the degree of treatment necessary. Treatment methods available to reduce total dissolved solids include: ion exchange, electrodialysis, reverse osmosis, and distillation.[3] Neutralization of the wastewater may be necessary to achieve proper pH. The treatment method employed at a centralized treatment facility will depend upon the characteristics of the industry's process waste streams with which the scrubber effluent is being combined.



Basis: 90 percent SO_2 removal

Figure 1. Sodium scrubbing process aqueous discharge rates.

Some industries (textile and paper mills) can use process waste streams containing sodium as a feed to the scrubber. The aqueous stream from the FGD system is then recombined with the industrial process waste streams and discharged to an on-site centralized waste treatment facility.[4] The treatment processes in such a centralized treatment facility vary with the specific industry. Typically, the treatment is designed to remove the dissolved and suspended solids and attain a neutral pH.

The adverse impacts of discharging aqueous scrubber wastes to the environment include potential degradation of the water quality (both surface and ground) of the receiving stream and the subsequent impact on users of that water. Improper treatment or disposal practices can result in aqueous wastes with high total dissolved solids being introduced into streams and aquifers that may serve as sources of water for other users.

Because of the problems associated with discharging wastes from the sodium scrubbing process in an environmentally acceptable manner, future applications of this technology will probably be limited to those facilities who can acceptably use evaporation ponds or who can use the aqueous waste stream as process make-up.

Solid Waste Impacts

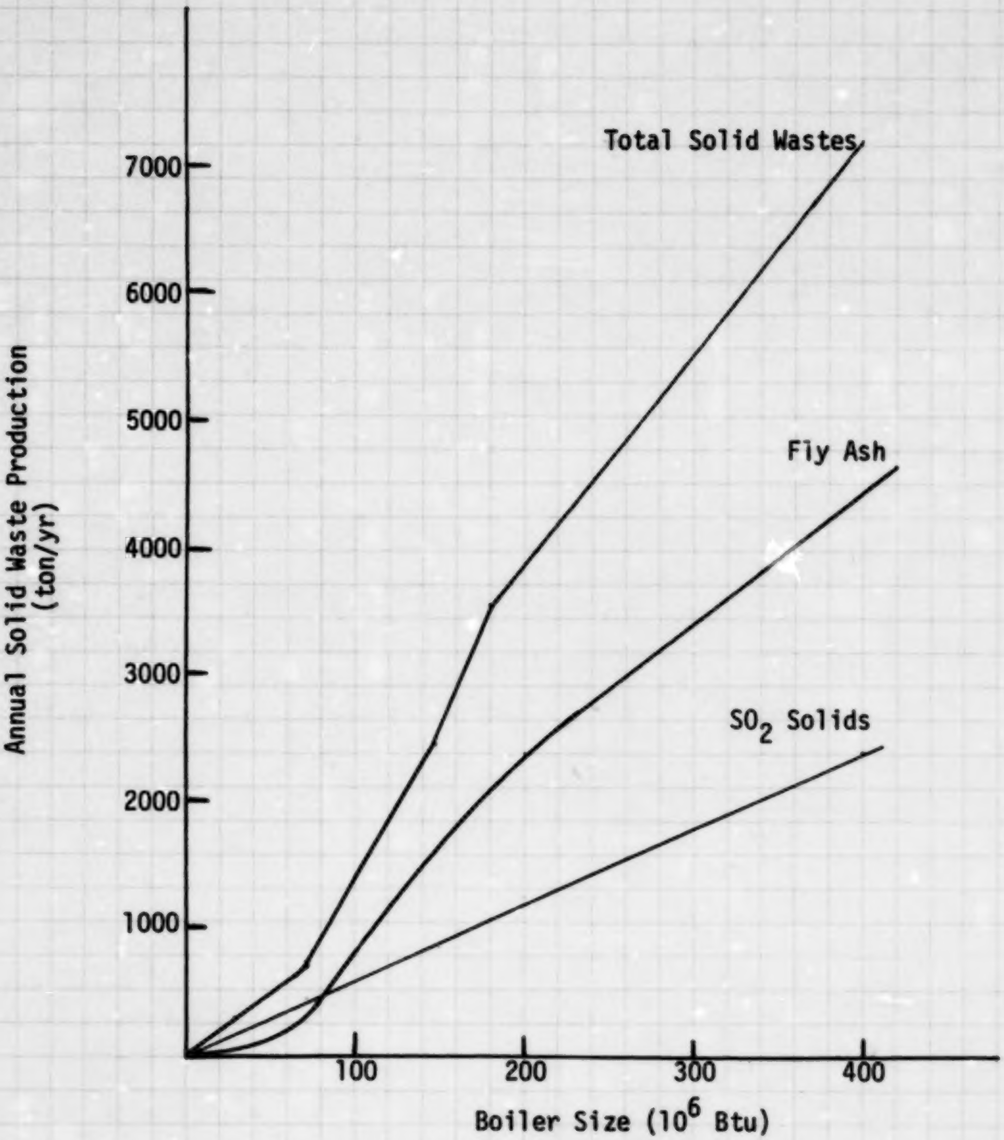
Solid wastes from industrial boiler FGD systems result from the dry solids produced in the spray drying process and the sludges produced in the limestone and DA processes. The dry solid waste product from the spray drying process will consist primarily of calcium or sodium salts, depending upon the type of alkali used as the SO_2 sorbent. Significant amounts of fly ash will also be present since the solids collection device associated with the spray dryer, probably a baghouse, will remove the particulates generated from the coal combustion process along with the spray drying solid wastes. Upstream particulate removal is not practical for this process since the spray dryer's performance is not adversely affected by the presence of fly ash[15] and dual particulate removal units would be unattractive from both an energy and economic view.

Waste sludges from dual alkali and limestone scrubbing systems are generally composed primarily of calcium sulfite/sulfate salts. Also present are dissolved trace elements (e.g., lead, arsenic and cadmium), which may contaminate the groundwaters and surface waters due to runoff and leaching from sludge disposal sites. The chemical composition and concentration of FGD sludge liquors vary with the different coal types used in industrial boilers. When a particulate collection device is not used upstream of the FGD system and the FGD system is being used to control both SO_2 and PM emissions, the trace element concentrations in the scrubber sludge are increased due to the addition of fly ash to the sludge.

Figures 2 and 3 show the quantities of solid wastes produced by the spray drying and DA FGD systems. Solid wastes for the spray drying system (Figure 2) are shown only for a low sulfur coal application and illustrate the contribution of fly ash to the overall solid wastes from the system. Solid wastes from a DA system are shown in Figure 3 for both high sulfur and low sulfur applications. Since the solid wastes produced in a limestone FGD system are similar in quantity and quality to those produced from a DA system, Figure 3 can also be used to approximate the wastes from a limestone system.

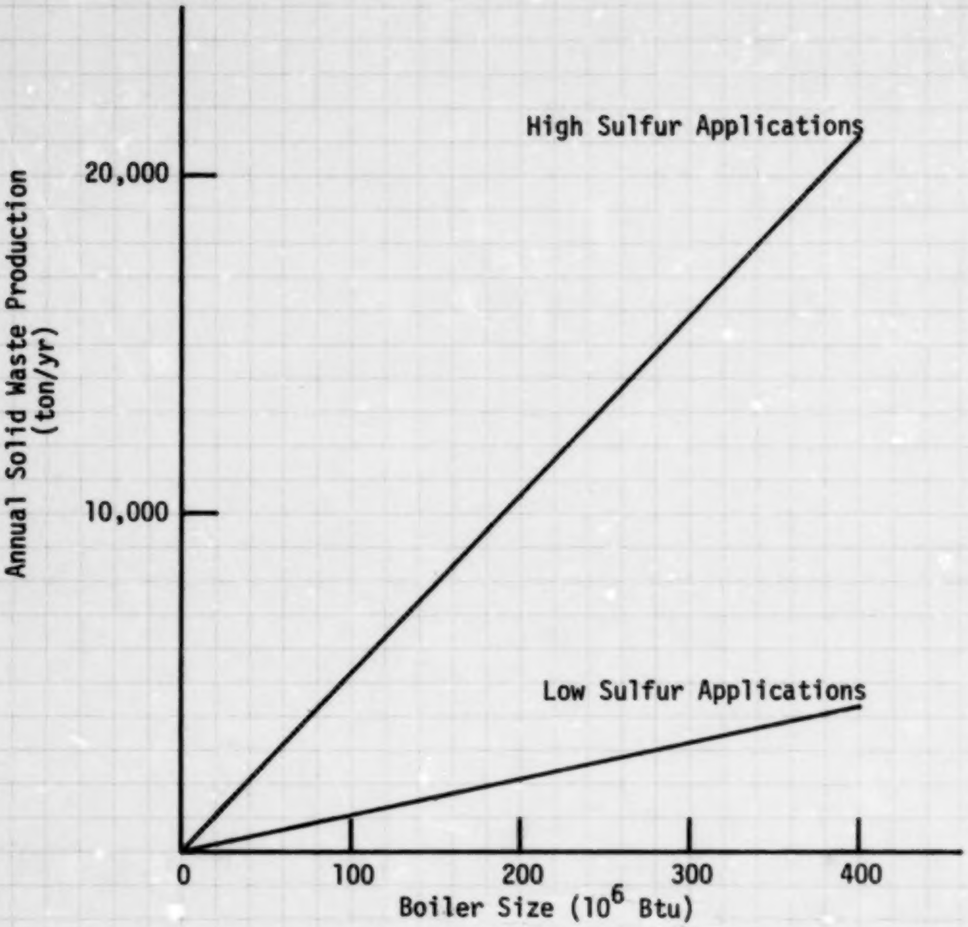
Solid wastes from spray dryers have physical properties similar to fly ash, and thus, can be handled in the same manner. Ponding and landfilling are currently the primary methods of disposing of collected fly ash. Off-site landfilling has been selected as the disposal method for the two operating spray drying systems operating at industrial boiler installations.

The main sludge disposal options for wet FGD systems include ponding and landfilling. Ponding is the simpler of the two methods, but is potentially the more harmful to the environment. Ponding involves slurrying the sludge to a pond, allowing it to settle and pumping the



Basis: 70% SO_2 Removal
Low Sulfur Coal
.03 #/10^6 Btu Particulate Emission

Figure 2. Spray drying process solid waste production.



Basis: 90% SO_2 Removal

Figure 3. Dual alkali process solid waste production.

Boiler Size (10^6 Btu/hr)

supernatant liquor either to a treatment process or back to the facility for reuse. Because there is always a hydraulic head on the waste in the bottom of the pond, the potential for leachates reaching ground-water sources beneath the pond is greater than for a landfill. Use of the pond area may be limited after disposal ceases, mainly because of the poor load bearing capabilities of the sludge compared to the original soil structure.[6]

Landfill disposal of FGD wastes in a specially prepared site requires some processing of the wet scrubber sludge (either stabilization or fixation) to obtain a soil-like material that may be loaded, transported and placed as fill. Stabilization refers to the addition of fly ash or other similar material to the sludge to produce only physical changes without any chemical reactions. Fixation is a type of stabilization which involves the addition of reagents (such as lime) to cause chemical reactions with the sludge. The objective of these treatment methods is to increase the load bearing capacity of the raw sludge and to decrease the permeability and correspondingly the mass transport rate of contaminants leaching out of the sludge.

At the present time the regulations governing solid waste disposal are not fully defined. EPA recently (May 2, 1980) issued Phase I final RCRA regulations covering the framework for management of solid wastes. Phase II regulations, those covering engineering design details of disposal sites will not be finalized until later. In addition, Congress is currently considering legislation that would exempt certain "special wastes" (as defined in the proposed regulations) from the possibility of being classified as hazardous until more data are gathered about their characteristics (two to three years).[7]

The Phase I RCRA regulations exempt fly ash, bottom ash, slag, and air pollutant emission control sludge produced in the combustion of fossil fuels from consideration as hazardous wastes. Therefore, since FGD solid wastes are currently exempt from hazardous waste regulations, they may be considered non-hazardous. Non-hazardous waste disposal

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management and techniques will be governed by Section 4004 of RCRA. This section requires states to implement disposal programs that will protect the environment (especially ground water) from contamination. EPA has also published Landfill Disposal of Solid Waste, Proposed Guidelines that will act as a guide to the states regarding the content of their disposal management programs.

Disposal of non-hazardous wastes will require at a minimum that a clay liner be used at the disposal site, that daily cover be applied, that access to the site be controlled, that ground-water quality at the site boundary be monitored, and that a final impermeable cover be placed and revegetation occur.[8] These activities are required, primarily, to protect ground water in the disposal area.

ENERGY IMPACTS

The energy requirements of each FGD process were also evaluated as a function of process size, fuel type, and level of SO_2 control. The major energy consumption area of the FGD systems was for electricity to operate the fans installed to overcome the pressure drop across the control systems. Energy for operating process pumps for wet FGD systems and rotary atomizers for spray drying systems were also considered. Energy requirements for stack gas reheat were not included in the process energy consumption totals.

The significant result of this analysis is that in all cases, FGD process energy requirements were shown to be about two percent of the net heat input to the boiler. The addition of stack gas reheat energy would increase the overall system energy requirements to about four to five percent of the net heat input to the boiler for the wet systems. The spray drying systems generally do not require reheat.

COST IMPACTS

Process costs (both capital and operating) were evaluated as a function of process size, fuel sulfur content, and SO_2 removal. The general approach used in developing the process costs consisted of four main steps. First, a series of material and energy balance calculations were performed for each process, to establish process stream flow rates and energy requirements as functions of unit size, SO_2 removal, and the amount of sulfur in the coal. Second, each of the FGD processes were divided into a number of process areas, or modules, which represented separate processing areas. Third, equipment sizes were then developed for each process module based on the results of the material and energy balances. Finally, capital cost estimates were prepared by contacting process equipment vendors for price quotations in the size range for the standard industrial boilers used in this study. All of the capital costs for each process area were developed in the form of direct capital costs which include all materials and labor installation costs. Except for the spray drying process which included a fabric filter, particulate control equipment costs were not included in this study.

Figures 4 and 5 show the capital costs for the industrial boiler FGD systems for high and low sulfur coal applications. As shown by these curves, the capital costs for the limestone and DA processes are essentially the same. The sodium scrubbing process is shown to have the lowest capital costs in all cases, which is probably the reason that over 75 percent of industrial boiler FGD systems are of this type. However, as mentioned previously, future applications of this process may be limited due to aqueous waste disposal problems. Finally, as shown in Figure 5, the spray drying process is the highest cost low sulfur coal FGD alternative in all cases. However, as mentioned above, costs for the spray drying process include costs for a fabric filter. If the fabric filter costs were eliminated, the spray drying process would become the lowest cost alternative. (Note that spray drying costs are based on 70 percent removal whereas wet scrubbing process costs are based on 90 percent removal. Previous calculations have shown that about a 10 percent capital cost reduction is realized in going from 90 to 70 percent removal for the wet systems).

Annualized costs of industrial boiler FGD systems are shown in Figures 6 and 7 for high sulfur and low sulfur coal applications. As shown by these curves, FGD system annualized costs exhibit the same

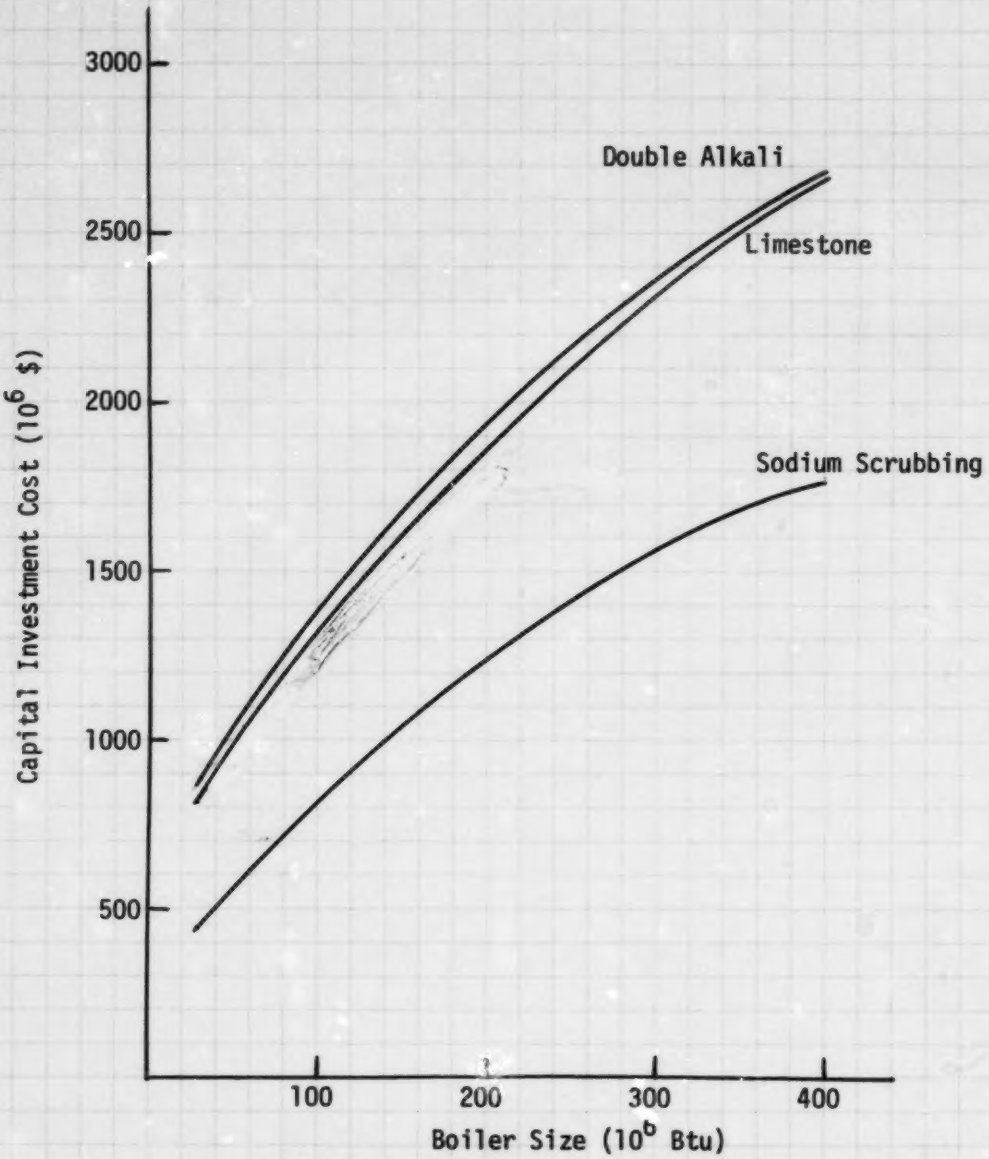

Figure 4. Capital Investment Costs - High Sulfur Coal Applications.

875

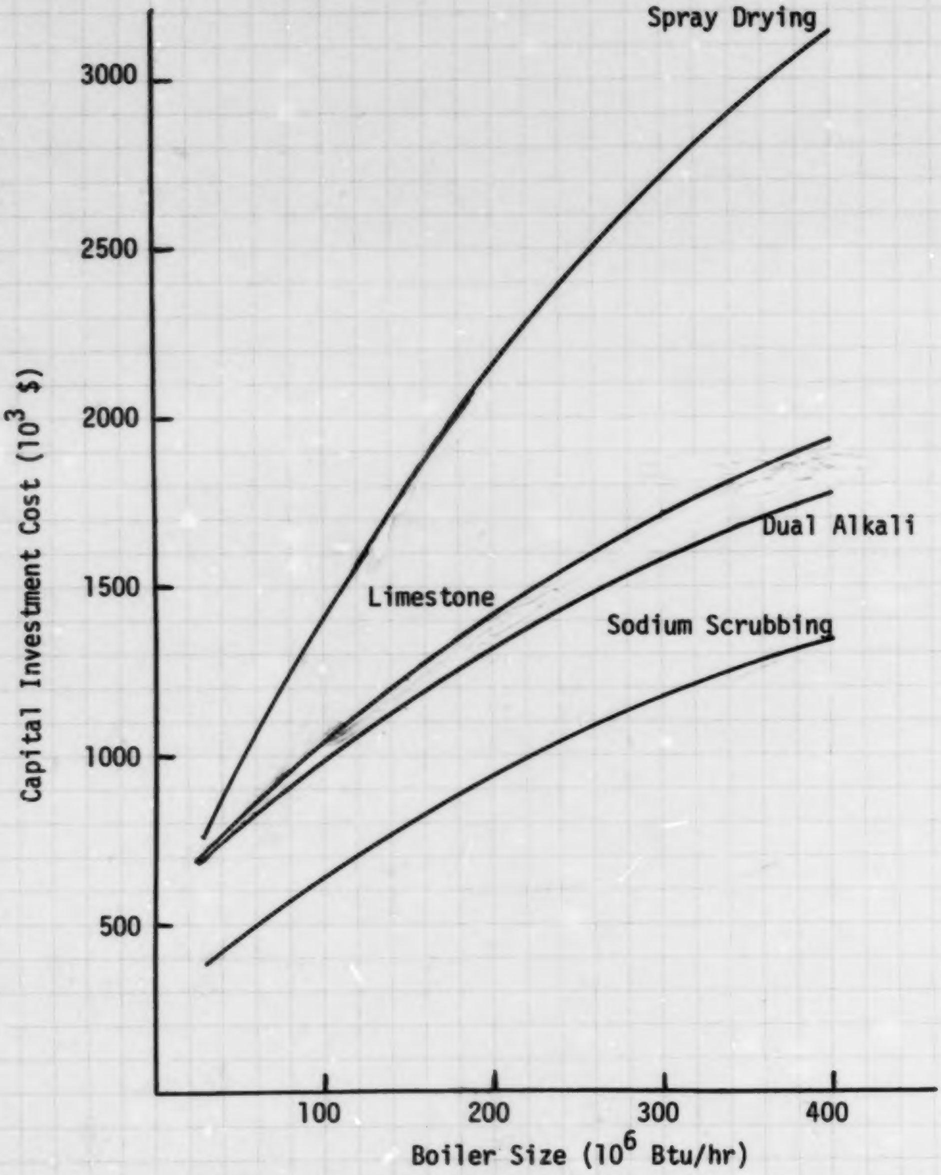

Figure 5. Capital investment cost - Low Sulfur Coal Applications.

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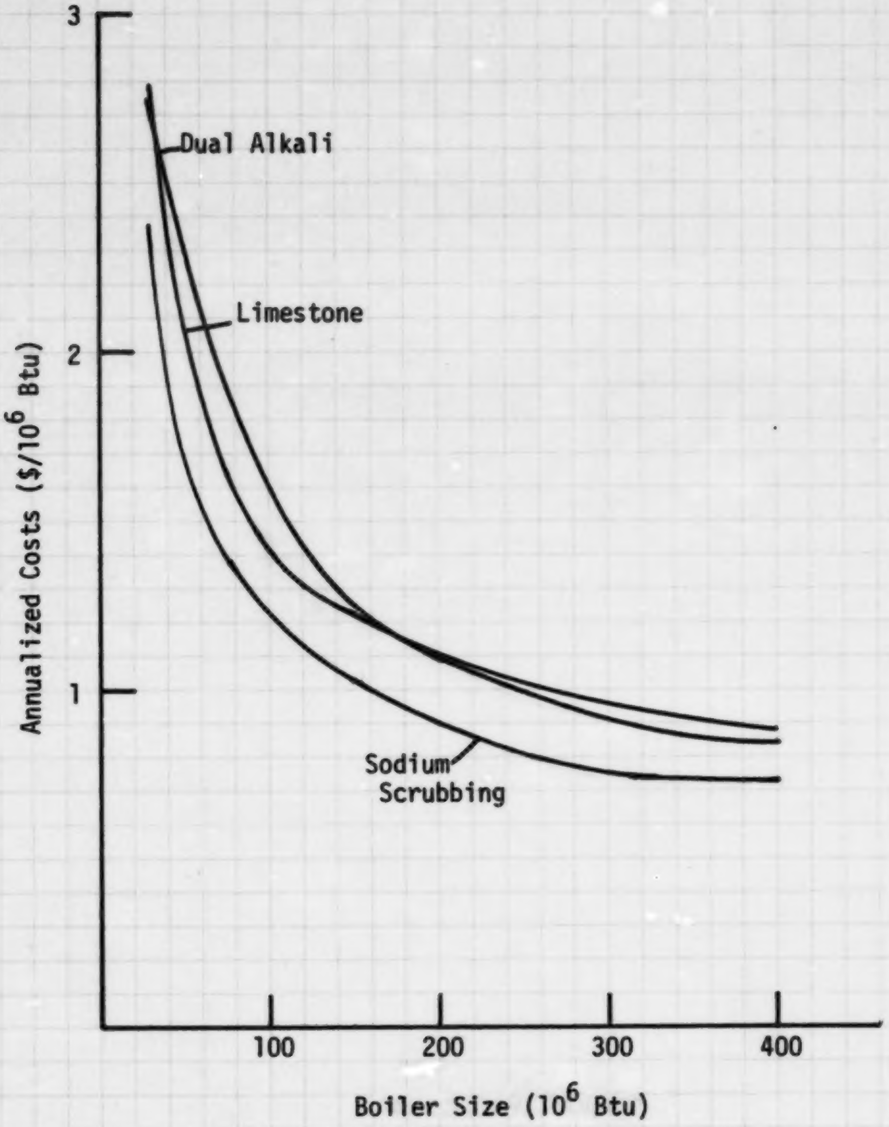

Figure 6. Annualized costs-High Sulfur Coal Applications.

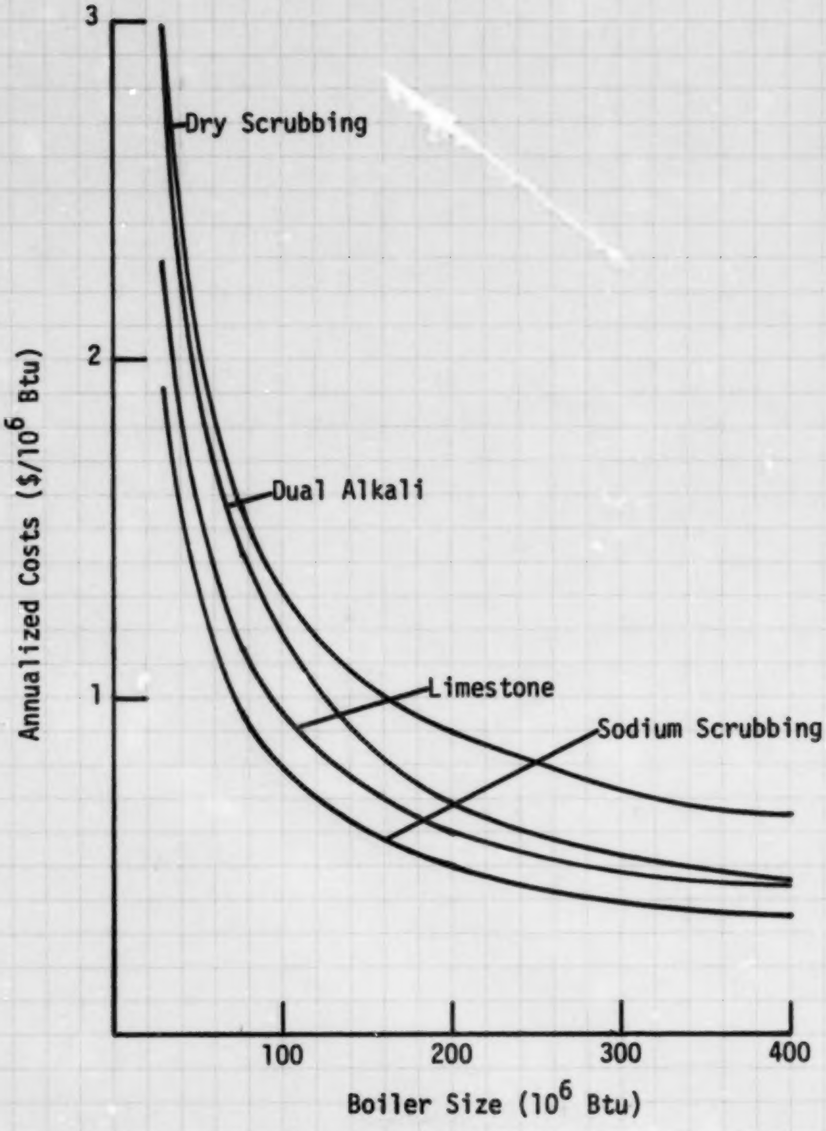

Figure 7. Annualized costs-Low Sulfur Coal Applications.

trends as do capital costs; that is the sodium scrubbing system costs are lowest in all cases, dual alkali and limestone costs are similar, and the spray drying costs are highest for the low sulfur coal cases. As before, the spray drying costs include the costs associated with particulate control using a fabric filter. Since the costs of the limestone and DA processes are so similar, and since the DA process makes up a significantly larger share of the market than the limestone process, the DA process will be used to represent costs of wet sludge processing processes in further cost comparisons.

The cost effectiveness of the various FGD processes was also determined. Cost effectiveness was defined as dollars per ton of removed SO_2 ($/ton SO_2) and was calculated by dividing the annualized process costs by the tons of SO_2 removed in a year assuming a 60 percent load factor. Results of these calculations show that both coal sulfur content and process size significantly affect the cost effectiveness of an FGD process. For a given size system, cost effectiveness increases with an increasing coal sulfur content. For a fixed coal sulfur content, cost effectiveness increases with increasing process size. Consequently, the most cost effective systems are those designed for large boilers burning a high sulfur coal, and the least cost effective systems are those designed for small boilers burning a low sulfur coal. Figure 8 illustrates these effects for the dual alkali processes. Curves developed for the other process showed similar effects.

An additional cost analysis was performed to estimate the overall impact on steam production costs of applying FGD control systems. The result of this analysis, shown in Figures 9 and 10 for high and low sulfur coal applications, was that FGD control systems would increase the steam production costs by up to 40 percent for small boiler installations. The steam production cost increase was on the order of 10 percent for larger sized boilers. Although these effects are shown only for the dual alkali process, other processes also showed higher steam production cost increases for the smaller boiler sizes.

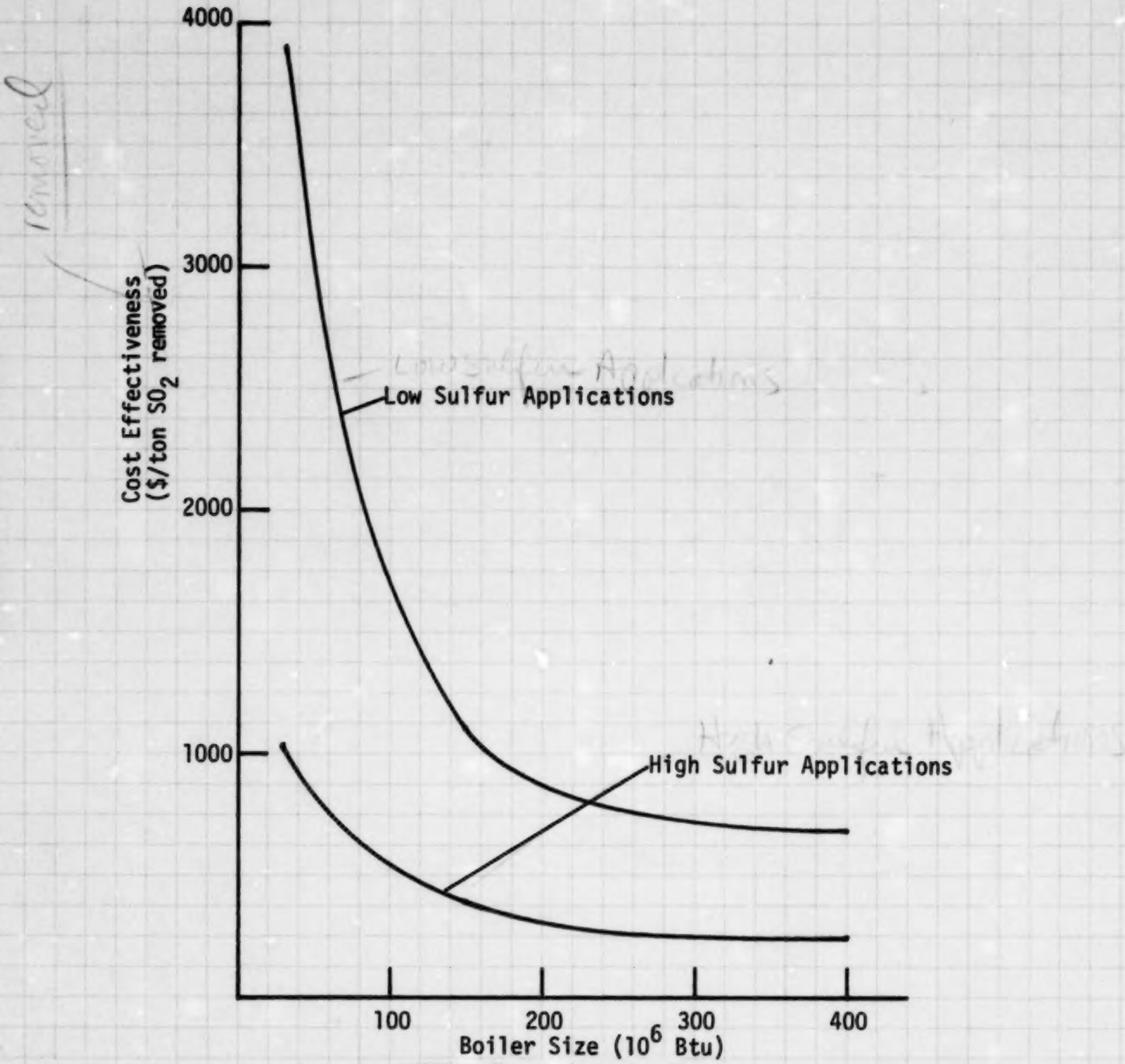

Figure 8. SO$_2$ removal cost effectiveness - Dual Alkali Process.

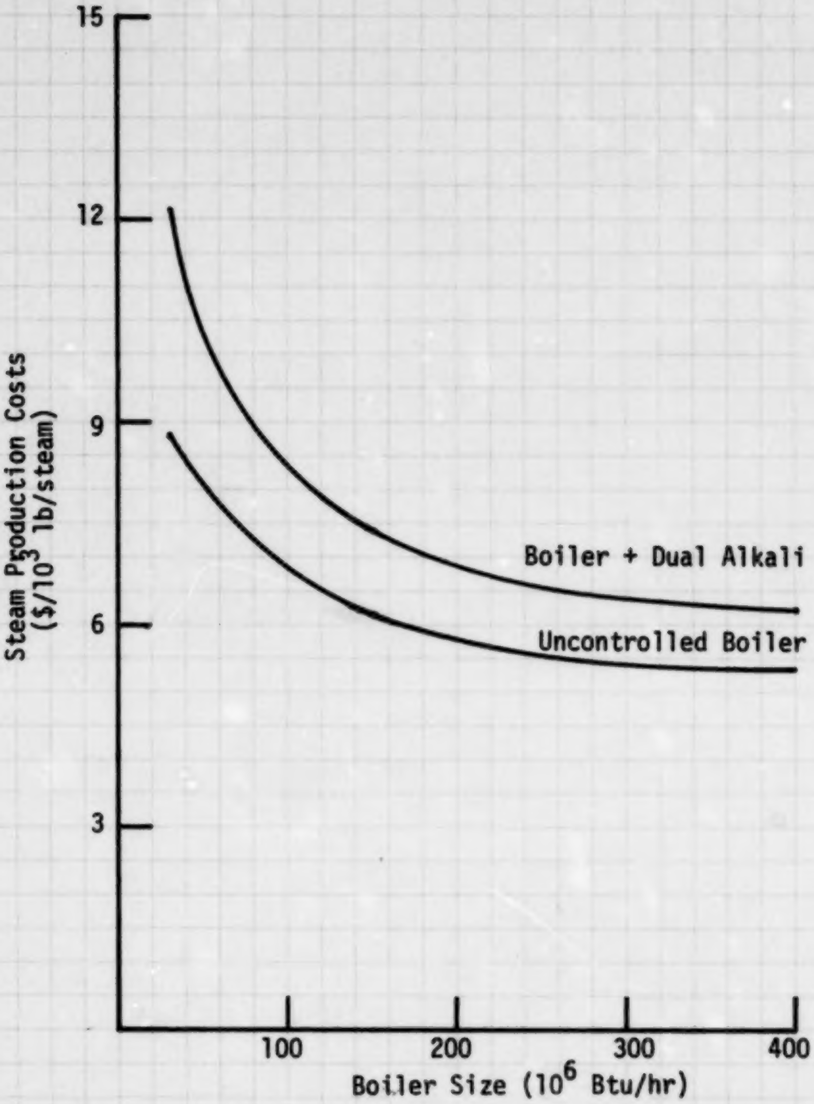

Figure 9. Increase in steam costs
Dual Alkali System
High Sulfur Coal.

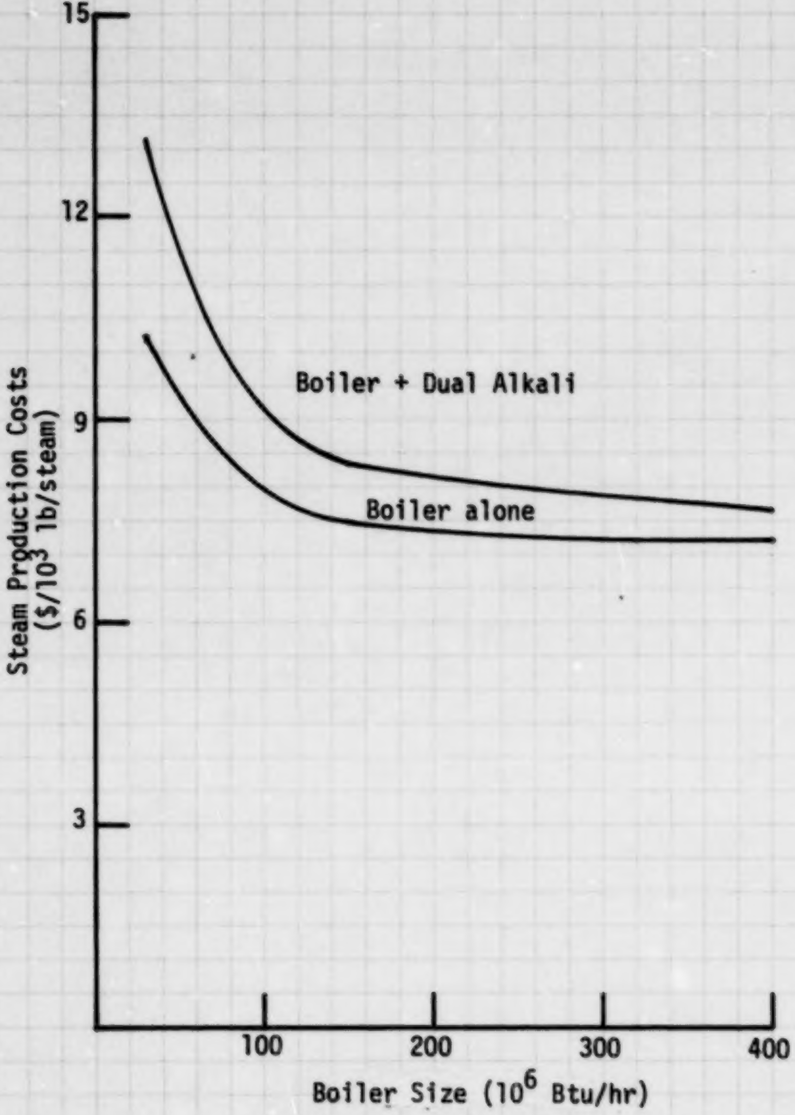

Figure 10. Increase in steam costs
Dual Alkali System
Low Sulfur Coal.

The results of the cost impact analyses showed that a clear variation in cost impacts exists between the large and small boilers. For all cost comparisons, the most severe impacts were shown to be for the smaller boilers and the least impact with the larger boilers.

SUMMARY

As mentioned previously, the purpose of performing an analysis of this type is to provide EPA with information that can be used to make decisions regarding the promulgation of new source performance standards for industrial boilers. Specifically, the results of these types of analyses can be used to establish technical limits for various control technologies, and to establish emission control levels as a function of boiler size that will equalize any adverse impacts across boiler sizes.

The FGD systems considered in this paper are all post-combustion flue gas cleaning processes and do not appear to be limited to any boiler type or size by technology considerations. This means that technological considerations would not prevent the application of FGD systems to even the smallest industrial boilers at SO_2 removal levels of up to 90 percent.

Although there do not appear to be technology limits that would prevent the application of FGD systems to small industrial boilers; examination of the environmental, energy, and cost impacts may show that application of FGD systems to small industrial boilers would result in greatly increased impacts. However, as previously presented, the environmental impacts are positive and the energy impacts are small for all applications. Consequently, neither of these criteria will provide a rationale for setting different control levels for different boiler sizes.

Cost impacts, however, vary considerably as a function of boiler size with the largest impacts being seen for the smaller sized boilers. This variation in FGD cost impacts between the small and large boiler sizes, may provide an incentive to EPA to promulgate different standards as a function of boiler size in order to equalize the cost impacts of a regulation. In fact, EPA currently has several cost impact analyses underway that will be used to help select appropriate emission control levels to equalize the cost impacts of applying FGD controls across the size ranges of industrial boilers. The results of these ongoing boiler specific analyses, along with consideration of other factors such as coal market penetration and distribution, will eventually be used to establish the level to which FGD systems will be required for industrial boiler applications.

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SULFUR DIOXIDE EMISSION DATA FOR AN INDUSTRIAL BOILER NEW SOURCE PERFORMANCE STANDARD

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ABSTRACT

To support development of standards of performance for new industrial boilers, the U.S. Environmental Protection Agency (EPA) initiated a test program on industrial flue gas desulfurization (FGD) applications to develop a data base for sulfur dioxide emissions control. The program included continuous monitoring of sulfur dioxide, oxygen, and moisture on both scrubber inlet and outlet. Systems selected included sodium scrubbing on oil and pulverized coal boilers, dual alkali systems on spreader stoker boilers, and a lime/limestone application on spreader stoker boilers. All boilers fired relatively high (2.5 to 4.0 percent) sulfur fuels.

Performance of the industrial FGD systems was generally superior to that previously examined in the utility sector, with respect to reliability and mean SO_2 removal. Variability of emissions from industrial FGD systems was somewhat less than for utility systems.

INTRODUCTION

In support of the effort by EPA to develop performance standards for industrial boilers, as detailed in the preceding Symposium paper, sulfur dioxide monitoring data were gathered at representative industrial FGD sites via continuous monitors. To determine representative test sites, the EPA Industrial Boiler FGD Survey (EPA-600/7-79-067b) of April 1979 was consulted. This report revealed that there were 126 operational FGD systems at 36 industrial sites in the United States. Of these 36 sites, 11 were on oil field applications and 12 on systems using waste process streams or plant products/by-products as the scrubber absorbent and/or co-firing non-fossil and fossil fuels in the boilers. Of the remaining

13 applications, 7 dual alkali, 4 sodium once-through, one lime, and one lime/limestone FGD system were identified.

Following presurveys to determine suitability for testing, four sites initially were chosen:

- ° Dual alkali - General Motors, Parma, Ohio
- ° Sodium/coal - General Motors, St. Louis, Missouri
- ° Sodium/oil - Mead Corporation, Stevenson, Alabama
- ° Lime/limestone - Rickenbacker AFB, Columbus, Ohio

DESCRIPTION OF SITES

GM-PARMA

The steam plant at the GMC Parma plant contains four boilers rated at a total combined steam generating capacity of 320,000 lb/h, two with a nominal capacity of 100,000 lb/h and two with a nominal capacity of 60,000 lb/h. Each is fired by a spreader stoker with traveling grates and operates with variable excess air rates in the 100 percent range. The larger boilers (1 and 2) are equipped with economizers that lower inlet flue gas temperatures to $\leq$275°F, and the smaller boilers (3 and 4) operate at an inlet temperature of $\leq$575°F. Each boiler is fitted with mechanical dust collectors for primary particulate control. Normal burning of medium- to high-sulfur (2 to 3 percent) eastern coal plus occasional firing of low-sulfur waste oil results in flue gas generally containing 500 to 1300 ppm SO_2 by volume; fuel analyses are presented in Tables 1 and 2.

FGD System

The FGD system consists of four double alkali scrubbing units that operated in a dilute mode. Figure 1 illustrates the process flow through each unit, and Table 3 lists design, operating, and performance characteristics of the system.

TABLE 1. ANALYSIS OF COAL AS RECEIVED - GM PARMA

Date	Ash, %	Moisture, %	Sulfur, %	Heating value, Btu/lb	Tons on hand	Tons received	Tons used
Dec. 1979	4.95	2.27	2.14	13,100	11,789	2,979	3,242
Jan. 1980	6.58	3.90	2.52	13,000	11,503	2,612	3,945
Feb. 1980	6.21	3.68	2.50	13,100	10,131	2,980	3,613
Mar. 1980	6.17	3.29	2.03	13,000	9,462	3,032	3,468
Apr. 1980	5.63	3.75	2.21	13,000	8,996	2,487	2,228

TABLE 2. ANALYSIS OF OIL AS RECEIVED - GM PARMA

Date	Moisture, %	Sulfur, %	Inert matter, %	Heating value Btu/gal	Gal used
Dec. 1979	1.625	0.85	6.0	148,350	42,100
Jan. 1980	3.0	0.9	6.4	147,300	52,950
Feb. 1980	2.1	0.8	3.9	145,050	57,500
Mar. 1980	3.3	0.6	7.8	150,810	44,200
Apr. 1980	16.0	0.53	0.82	149,500	33,800

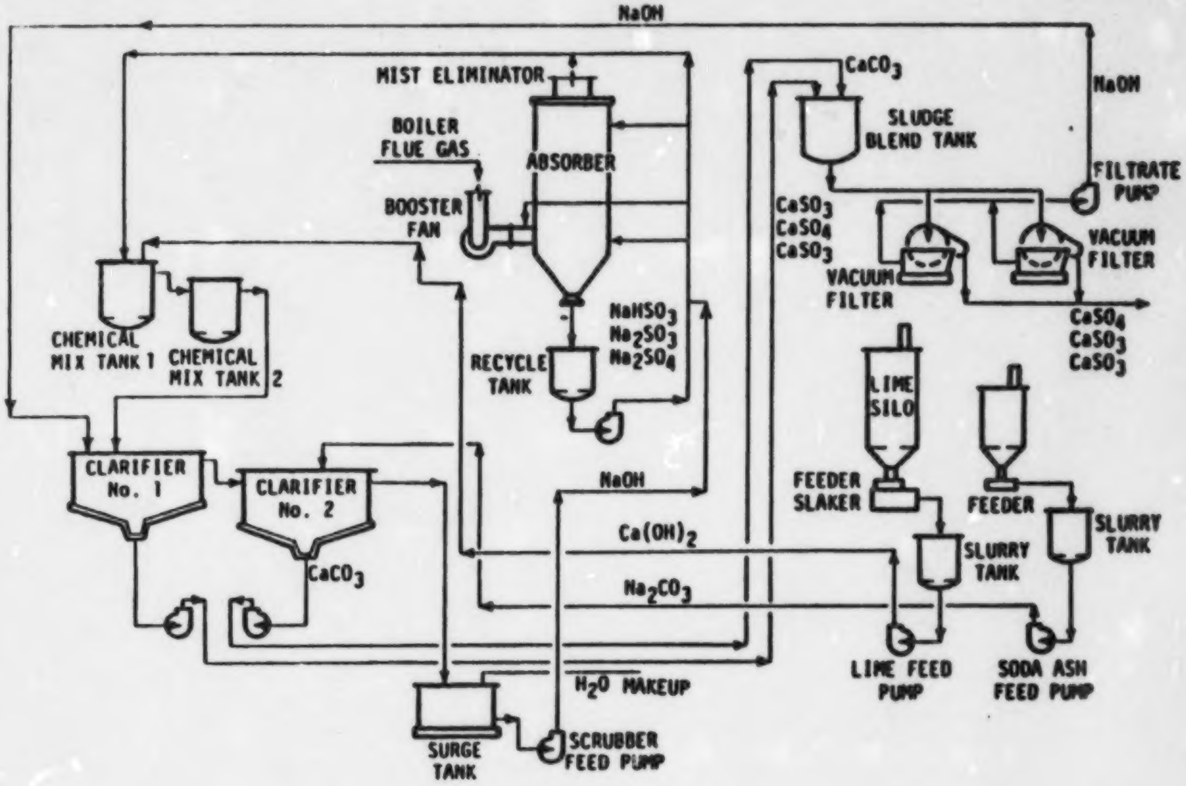
NaOH
MIST ELIMINATOR
BOILER FLUE GAS
ABSORBER
BOOSTER FAN
CaCO$_3$
SLUDGE BLEND TANK
NaOH
FILTRATE PUMP
CaSO$_3$
CaSO$_4$
CaSO$_3$
VACUUM FILTER
VACUUM FILTER
CaSO$_4$
CaSO$_3$
CaSO$_3$
CHEMICAL MIX TANK 1
CHEMICAL MIX TANK 2
RECYCLE TANK
NaHSO$_3$
Na$_2$SO$_3$
Na$_2$SO$_4$
LIME SILO
FEEDER SLAKER
FEEDER
SLURRY TANK
SLURRY TANK
CLARIFIER No. 1
CLARIFIER No. 2
CaCO$_3$
NaOH
Ca(OH)$_2$
Na$_2$CO$_3$
H$_2$O MAKEUP
LIME FEED PUMP
SODA ASH FEED PUMP
SURGE TANK
SCRUBBER FEED PUMP

Figure 1. Process flow diagram of a double alkali scrubbing unit.

TABLE 3. DESIGN, OPERATING, AND PERFORMANCE CHARACTERISTICS OF DOUBLE ALKALI SCRUBBING SYSTEM

Application	Four coal-fired stoker boilers
Fuel characteristics	Coal: 25,600 to 31,400 J/g (11,000 to 13,500 Btu/lb), 1.5 to 3.0 percent sulfur (waste oil also burned, but in small quantities compared with coal)
System design	Four three-stage multiventuri flexitry scrubber modules
Process mode	Dilute active alkali
Pressure drop	25 to 33 cm H_2O (10 to 13 in. H_2O)
Status	Operational
Startup date	March 1974
Inlet gas conditions	
Flow rate	30.9 m^3/s at 27°C (65,500 acfm at 80°F)
SO_2	800 to 1300 ppm
O_2	Not available
Particulate	0.7 g/m^3 (0.3 gr/scf)(dry)
SO_2 outlet concentration	20 to 130 ppm
SO_2 removal efficiency	90 to 99 percent
Scrubbing solution characteristics:	
pH	5.5 to 7.5
Total sodium	0.58 to 0.96 molar
Active sodium	0.087 to 0.13 molar
Calcium ion	305 to 490 ppm
Soda ash makeup	>0.1[a] mole/mole SO_2 removed
Lime utilization	1.32 to 1.90 mole/mole S in cake
Filter cake solids	40 to 55 percent
Filter cake disposal	Offsite landfill

[a] Nonsteady-state operations result in makeup rates of 0.028 to 0.05 mole NO_2 per mole SO_2 removed.

Flue gases enter through a prequench section at the bottom of each unit and then flow in countercurrent direction to an aqueous sodium hydroxide sulfite-bisulfite solution. Each is a two-tray, impingement-type unit with feed and recycle streams entering at the top. The absorption trays (Koch) have movable self-adjusting bubble caps that respond to variations in gas flow. Pressure drop through each unit is designed at 19 cm (7.5 in.) H_2O, and the maximum liquid-to-gas ratio is 2.7 liters/m^3 (20 gal/100 ft^3). The liquid feed is composed of about 20 percent fresh feed and 80 percent recycle. For control of entrained liquor, each unit is equipped with a high-efficiency mist eliminator. Acting as an impingement separator, the mist eliminator is composed of corrugated profile plates assembled with phase separating chambers.

Process Control

In the normal mode of operation, a sidestream from the reagent recirculation loop in the FGD absorber section is constantly fed to the chemical mix tank, where calcium carbonate ($CaCO_3$) slurry is fed for regeneration of caustic. Overflow from this tank enters another mix tank for further reaction.

The regeneration solution, with a high concentration of fly ash and calcium precipitates, flows to two reactor clarifiers in series. In the first clarifier, the solution goes through additional reaction and solids separation. Liquid effluent is then pumped to the second clarifier, where it is softened by the addition of sodium carbonate (Na_2CO_3). Solution from this tank is recycled to the scrubber recirculation loop. Underflow from both clarifiers is pumped to the sludge blend tank for batch processing through a vacuum filter, and cake from the filter is hauled to a landfill for disposal. Filtrate is returned to the primary clarifier for recovery of sodium hydroxide (NaOH).

The chemical reactions that take place in the system are:

SO_2 absorption: $SO_2 + H_2O \; T \; HSO_3^- + H^+$

$NaOH + HSO_3^- + H^+ \rightarrow 2Na^+ + SO_3^= + 2H_2O$

$SO_3^= + SO_2 + H_2O \rightarrow 2HSO_3^-$

Regeneration: $Na_2SO_4 + Ca(OH)_2 \rightarrow CaSO_4\downarrow + 2NaOH$
(chemical mix tank) $Na_2SO_3 + Ca(OH)_2 \rightarrow CaSO_3\downarrow + 2NaOH$

Softening: $Na_2CO_3 + Ca^{++} \rightarrow 2Na^{++} + CaCO_3\downarrow$
(Clarifier 2)

GM-ST. LOUIS

The power plant at the General Motors Assembly Division facility located in St. Louis, Missouri contains four coal-fired steam generators. The exhaust from any of four of the steam generators can be passed through two Peabody scrubbers. For this project, only the exhaust gases from Boiler No. 4, a Babcock & Wilcox (B&W) pulverized coal-fired unit, were passed through the FGD system. A schematic of the entire system is shown in Figure 2.

Throughout the sampling program, the coal fired in the boiler was supplied by the Peabody Coal Company, Illinois Division, River King Mine. It was sub-bituminous, with an ash content of 10 percent and a sulfur content between 3.25 and 3.73 percent. The coal was delivered to the facility by truck. Proximate and ultimate analyses of the coal shipped was supplied by the Peabody Coal Company. A typical analysis for one shipment is given in Table 4.

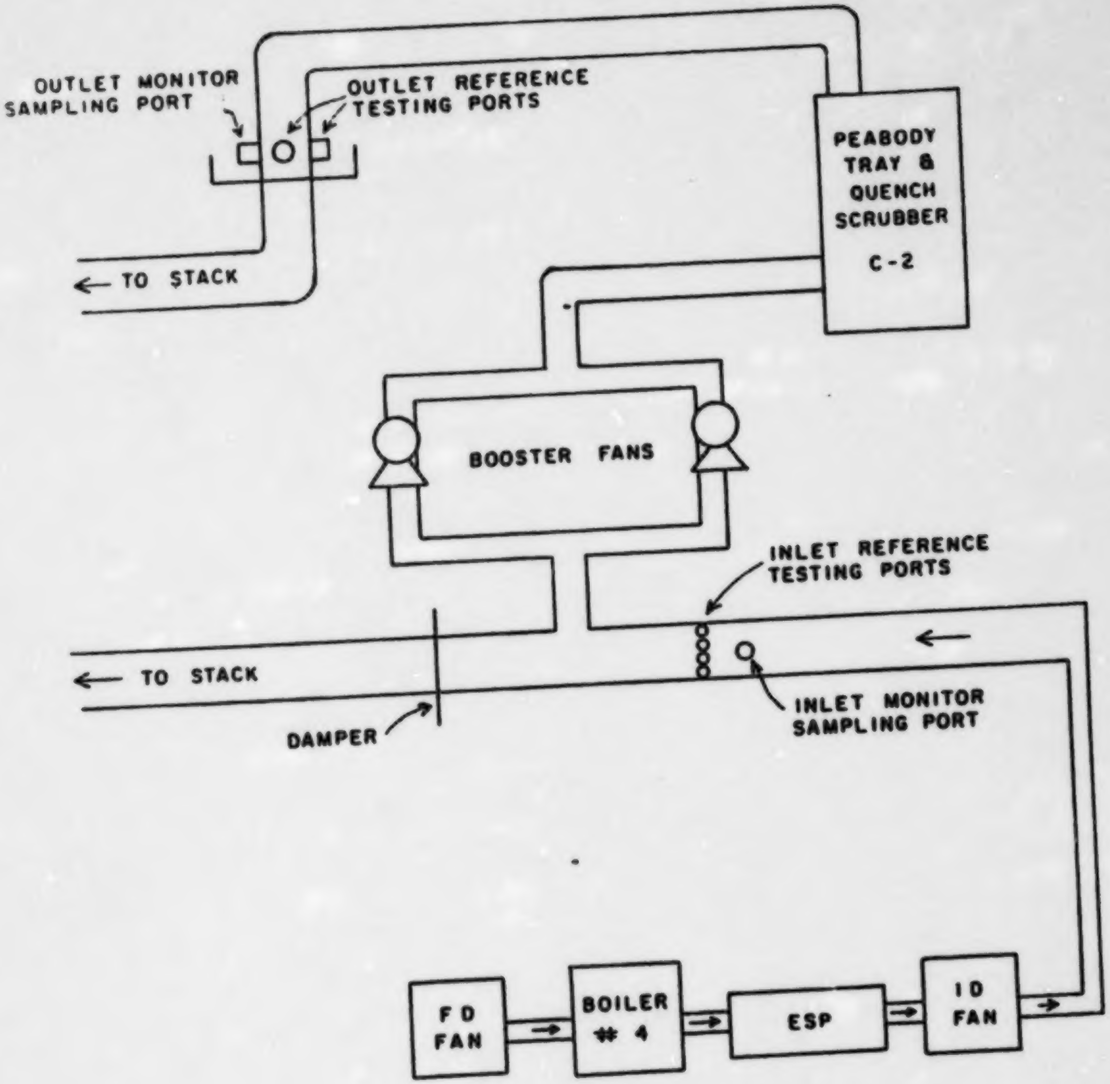
OUTLET MONITOR SAMPLING PORT
OUTLET REFERENCE TESTING PORTS
← TO STACK
PEABODY TRAY & QUENCH SCRUBBER
C-2
BOOSTER FANS
INLET REFERENCE TESTING PORTS
← TO STACK
DAMPER
INLET MONITOR SAMPLING PORT
FD FAN
BOILER # 4
ESP
ID FAN

Figure 2. Schematic of FGD system - GM-St. Louis

TABLE 4. TYPICAL COAL ANALYSIS AT GM ST. LOUIS

PROXIMATE ANALYSIS	RIVER KING MINE
As Received	Truck
Moisture	12.5%
Ash	10.4%
Volatile	35.6%
Fixed Carbon	41.5%
	100.0
BTU (± 100)	11,000
BTU (Dry)	12,571
Sulfur	3.1%
ULTIMATE ANALYSIS	
Dry Basis	
Carbon	72.0%
Hydrogen	5.1%
Nitrogen	1.3%
Chlorine	0.1%
Sulfur	3.5%
Ash	11.9%
Oxygen (Diff.)	6.1%
	100.0
Ash Fusion (Reducing)	2110 F
Grindability	55

The maximum steam generation for Boiler No. 4 is 150,000 lb/hr. The exhausted gases pass through mechanical collectors and an electrostatic precipitator for removal of particulate matter prior to entering the FGD system.

Flue Gas Desulfurization System

The FGD system is a Peabody tray and quench liquid scrubber. It consists of a three-stage impingement tower with a chevron mist eliminator. The scrubbing medium is a 50 percent aqueous sodium hydroxide (NaOH) solution. The liquid loop is open with a 35 gal/min makeup. The liquid waste is discharged to the city sewer system after the Na_2SO_3 has been oxidized to Na_2SO_4 and the pH adjusted to neutrality.

Figure 2 is a schematic of the FGD system at GM-St. Louis.

The sulfur dioxide removal efficiency is rated at 90 percent with an inlet SO_2 concentration of 2000 ppm.

The exhaust gases are reheated via steam coils surrounding the duct.

MEAD CORPORATION[3]

The Mead Paperboard Company in Stevenson, Alabama is an integrated neutral sulfite semi-chemical (NSSC) pulp and paperboard mill producing corrugating medium from 635 T/D NSSC pulp and recycled fiber. The mill has an SCA-Billerud chemical recovery system to recover sodium and sulfur cooking chemicals.

Two Combustion Engineering type 28-VP-14W oil-fired package boilers are employed to fill steam generation demands not met with a "base loaded" recovery boiler and hog fuel-fired steam generator.

The package boilers are fired with No. 6 fuel oil containing 4 percent sulfur with a gross calorific value (GVC) of 39,929 KJ/KG (17,167 BTU/lb). These units operate in tandem and each produces a maximum of 170,000 pounds of steam per hour. The fuel is received by barge and stored in a 4,000,000 gallon intermediate storage and transferred to a 3,000,000 gallon tank. During the normal operating conditions, fluctuations in the steam demand may be as great as 40,000 lb/hour.

Table 5 shows a typical fuel analysis.

FGD System

The flue gas desulfurization unit is an AirPol Venturi scrubber, which started up in 1975.

The sulfur dioxide removal system was designed to emit 125 lbs/hr at an inlet loading of 1250 lbs per hour. Particulate removal efficiency is rated at 80 percent.

Flue gases discharged from the oil-fired power boilers proceed through ducting and enter the AirPol circular type venturi scrubber. The scrubber operates on the principle of thorough atomization of the scrubbing liquid. The scrubbing liquid enters the scrubber through a large open pipe and is sprayed around the entire inlet creating a complete wet flooded approach.

The scrubbed gases and liquid with collected solid particles continue to an AirPol spin-vane separator which further cleans the gas to remove particulate and SO_2. Finally, the gas spins through separator spin vanes exiting at the top of the unit, while liquid and solids exit at the bottom of the vessel to a separate recycle tank for recirculation to the scrubber. The gas exits through a fiberglass stack.

Scrubber effluent is purged at a 15-30 gpm rate and used as process liquor. Makeup alkalinity is sodium carbonate/sodium hydroxide at 1.3 lbs

TABLE 5. TYPICAL FUEL OIL ANALYSIS AT MEAD

ASH	%	0.02
CARBON	%	88.12
DENSITY	lbs/gal	9.0825
HEATING VALUE	Btu/lb	17,167
HEATING VALUE	Btu/gal	155,919
HYDROGEN	%	7.08
NITROGEN	%	0.14
OXYGEN	%	0.58
SULFUR	%	4.06

per gallon. Evaporation is compensated by fresh water makeup. Scrubber pH is controlled to 7.0-8.5.

Figure 3 is a representative schematic of the Mead FGD system.

Process Control

The FGD system at Mead has essentially no process control other than periodic checks of liquor pH and carbonates from grab samples. During the test period, manpower commitments to other plant operations and plant emergencies resulted in infrequent pH checks and several excursions. For example, on November 6, 1979, the scrubber pH was below 7.0 for the entire 24-hr period and SO_2 removal averaged 49.0 percent. By November 10, 1979, pH ranged from 7.7 to 12.2 and SO_2 removal averaged 98.6 percent.

RICKENBACKER AFB[4]

The boiler system at Rickenbacker Air Force Base consists of six industrial-size boilers designed to supply high-temperature water for building and water heating at the base. The five older boilers are rated at 30 x 10^6 Btu/h, and the newer boiler is rated at 60 x 10^6 Btu/h. The boilers are stoker fired, with coal spread on under-air-flow perforated grates and burned at a typical ash bed depth of 3 to 5 inches. Grates are mechanically rotated in all but the oldest operating boiler to remove ash to the pneumatic disposal system. Ash must be removed manually from the grate of the oldest unit in operation. Analyses of coal burned during the test period are summarized in Table 6.

FGD System

The FGD system consists of a mechanical collector, Swedish Bahco scrubber tower, lime storage and handling system, clarifier (thickener), booster fan, sludge disposal pond, and associated duct work, pumps, and controls.

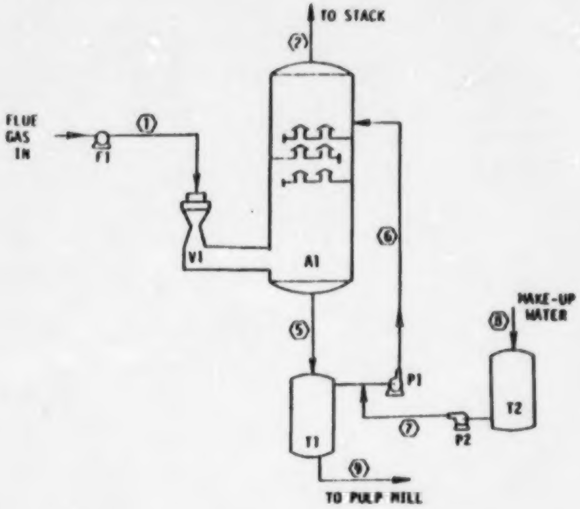
TO STACK
⑦
FLUE GAS IN
F1
①
V1
A1
⑥
⑤
MAKE-UP WATER
⑧
P1
T2
T1
⑦
P2
⑨
TO PULP MILL

Figure 3. Mead Paperboard Co.
FGD System

EQUIPMENT LIST

EQ. NO.	NAME	SIZE, FEET	MATERIAL	NO.
A1	ABSORBER		316 SS	1
F1	F.D. FAN		C.S.	1
P1	RECIRCULATION PUMP			1
P2	CAUSTIC PUMP			1
T1	RECIRCULATION TANK		FIBERGLASS	1
T2	SODIUM CARBONATE TANK			1

GAS STREAM NO.	①	②	③	④
RATE, lb/hr x 1000				
acfm x 1000	175	115		
PARTICULATES, lb/hr				
SO_2, lb/hr				
TEMPERATURE, °F	450	140		
H_2O, lb/hr				
O_2, %				
NO_x, lb/hr				
VELOCITY, ft/sec				

LIQUID STREAM NO.	⑤	⑥	⑦	⑧	⑨	⑩	⑪	⑫	⑬	⑭	⑮	⑯	⑰
RATE, lb/hr x 100													
gpm	3000	600	20-25	20-25									
TOTAL SOLIDS, %													
pH			7.0-8.5										
TEMPERATURE, °F													
H_2O, lb/hr													
SULFITE, mg/l													
SULFATE, mg/l													
CHLORIDE, mg/l													
SPECIFIC GRAVITY	1.1-1.2												

911

TABLE 6. COAL DELIVERED AND BURNED AT RICKENBACKER AFB

Date delivered	Sulfur content (dry), %	Heating value (dry, Btu/lb	SO_2 emission rate, lb/10^6 Btu	Date burned in 1980
1/16-1/17	2.9	13,080	4.21	Stockpiled
1/17-1/29	3.0	13,140	4.34	Stockpiled
2/7-2/15	3.9	13,540	5.47	2/11-2/19
2/15-2/28	3.7	13,470	5.22	2/19-3/3
2/29-3/6	2.3	12,900	3.39	3/3-3/10
3/6-3/14	3.1	13,260	4.44	3/10-3/19

Untreated flue gas from the individual boilers enters a common header equipped with a bypass stack and is fed through a mechanical collector where primary particulate removal takes place. The mechanical collector has a design removal efficiency of 70 percent and was installed primarily to reduce wear on the booster fan, which is located immediately downstream. The fan introduces the partially cleaned flue gas into the scrubbing tower where SO_2 removal takes place.

The scrubber is a vertical tower consisting of two inverted venturi scrubbing stages. Untreated gas is introduced into the first stage, where it is diverted downward to impinge on the liquid slurry surface of the mill. The gas then rises through the first stage venturi, where it intimately mixes with the slurry droplets now entrained in it. The partially scrubbed gas is then diverted downward onto the liquid slurry surface in the second stage pan, and the process is repeated. The treated gas is then directed upward into a cyclonic mist eliminator where entrained slurry droplets are removed, and then emitted through a stub stack to the atmosphere.

The reagent for the scrubbing system is fine mesh limestone or pebble lime slurry. Normally limestone is used, but for the purpose of this test, lime was employed. The reagent slurry is introduced through the scrubbing system in a countercurrent fashion.

The spent scrubbing solution is discharged to the thickener where waste solids settle out. Thickener overflow is returned to the mixing tank. Underflow from the thickener is discharged to a 5-acre Hypalon-lined disposal pond located approximately 400 feet from the FGD system. Figure 4 is a flow diagram of the Research-Cottrell (R-C)/Bahco scrubbing system.

Process Control During Test

During the test, the boiler system was operated in normal fashion. Average total boiler load during data collection days was 109 x 10^6 Btu/h,

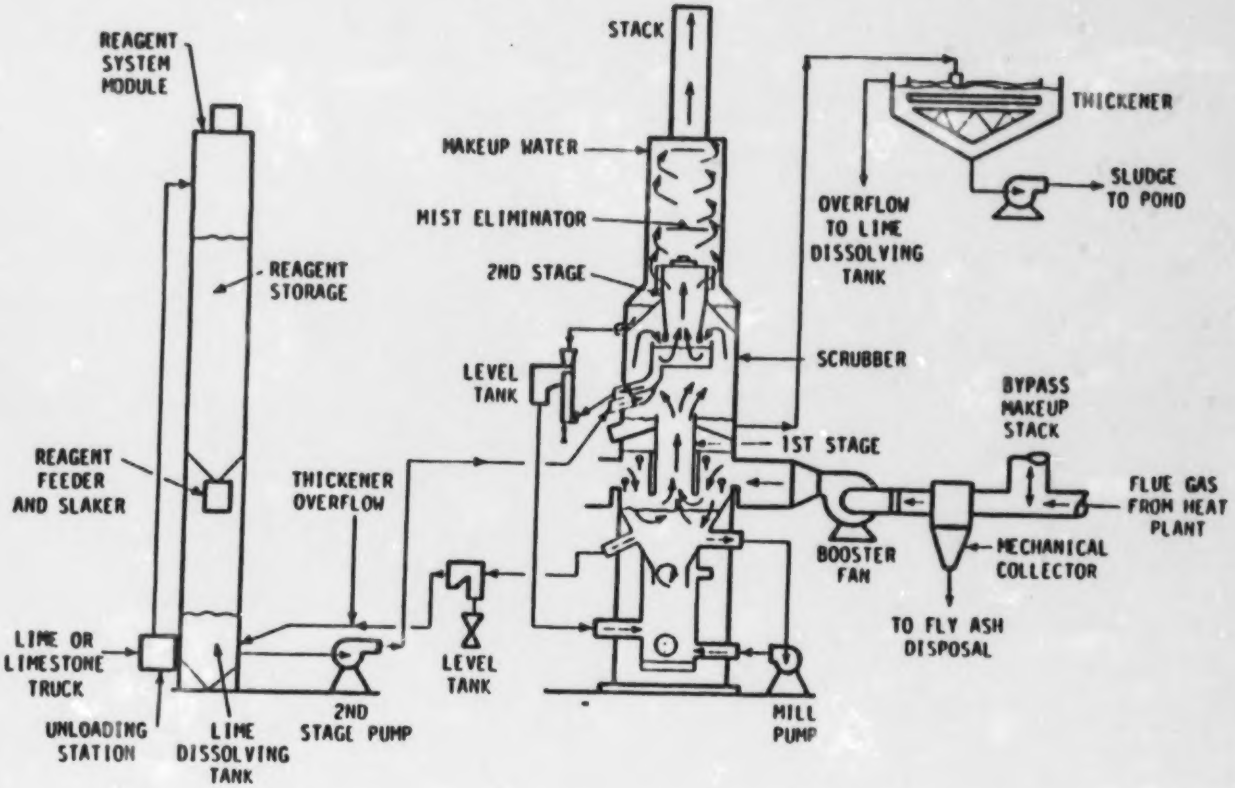
REAGENT SYSTEM MODULE
REAGENT STORAGE
REAGENT FEEDER AND SLAKER
LIME OR LIMESTONE TRUCK
UNLOADING STATION
LIME DISSOLVING TANK
THICKENER OVERFLOW
2ND STAGE PUMP
LEVEL TANK
STACK
MAKEUP WATER
MIST ELIMINATOR
2ND STAGE
LEVEL TANK
SCRUBBER
1ST STAGE
MILL PUMP
OVERFLOW TO LIME DISSOLVING TANK
THICKENER
SLUDGE TO POND
BYPASS MAKEUP STACK
FLUE GAS FROM HEAT PLANT
BOOSTER FAN
MECHANICAL COLLECTOR
TO FLY ASH DISPOSAL

Figure 4. R-C/Bahco scrubbing system flow diagram.

which is 52 percent of the total system capacity. Generally, three to five boilers were sufficient to meet heat demand during the period, with the remainder on standby status.

During the test period, the lime feed rate from storage to the slaker was the only significant control needed to achieve desired SO_2 removal efficiency. Plant operators on all shifts made periodic checks of the FGD performance, as evidenced by the continuous emission monitor (CEM) output, and adjusted the lime feed rate as needed. Because the CEM used a measuring range ratio of 10:1 for inlet and outlet SO_2 readings and recorded them on the same chart, it was easy for plant operators to note when adjustments were needed. The only significant periods recorded during which the FGD system did not operate at approximately 90 percent or better SO_2 removal occurred when the lime feed rate control was lost because of mechanical breakdown of the lime slaker.

MONITORING SYSTEM DESCRIPTION

At each site, the continuous SO_2 monitoring system consisted of an SO_2 monitor with an oxygen (O_2) monitor mounted at the point where sample lines enter the SO_2 monitor. Although site-to-site deployment of equipment varies, Figures 5 and 6, as used at the GM-Parma site, are representative of a continuous SO_2 monitoring system.[5]

Sample gas is pulled through a stainless steel mesh filter screen inside the ductwork, through a probe equipped through a blowback system, and into a sample line. Because of the high moisture content of FGD outlet gases, a conditioning system is normally required for accurate moisture determination in outlet emission calculations. This system consists of 1/4-inch stainless steel tubing coiled and immersed in a thermostatically controlled water bath at 10°C (50°F). No conditioning system is normally required for FGD inlet gases.

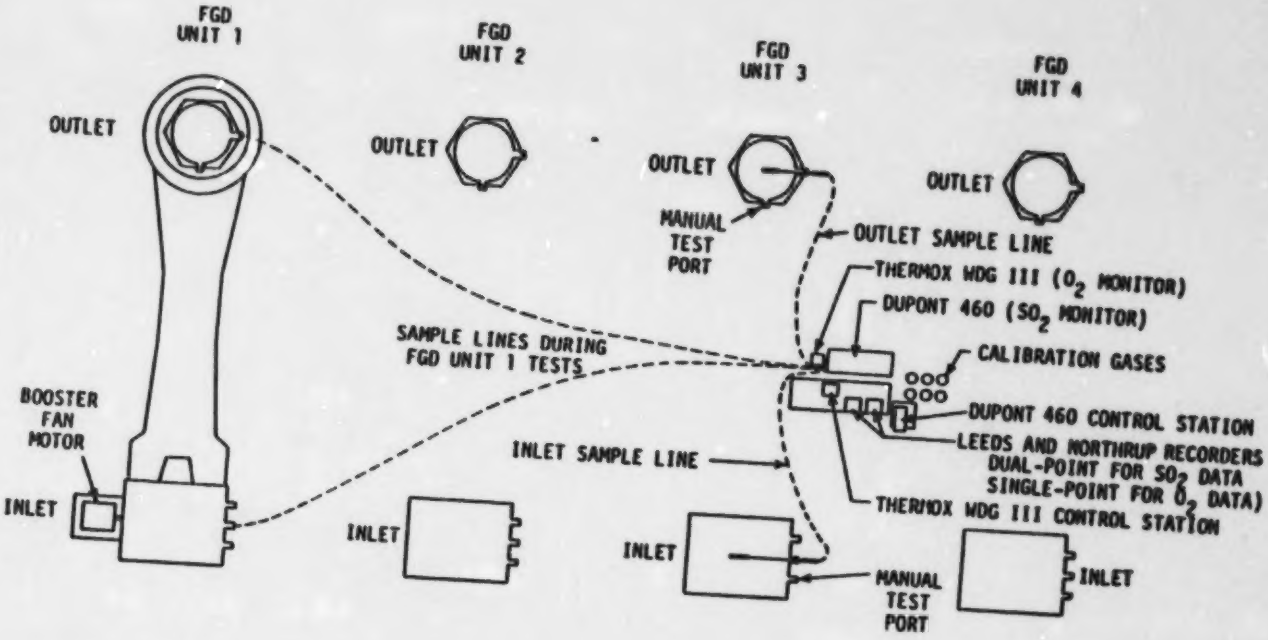
FGD UNIT 1
FGD UNIT 2
FGD UNIT 3
FGD UNIT 4
OUTLET
OUTLET
OUTLET
OUTLET
MANUAL TEST PORT
OUTLET SAMPLE LINE
THERMOX WDG III (O_2 MONITOR)
DUPONT 460 (SO_2 MONITOR)
CALIBRATION GASES
SAMPLE LINES DURING FGD UNIT 1 TESTS
DUPONT 460 CONTROL STATION
LEEDS AND NORTHRUP RECORDERS DUAL-POINT FOR SO_2 DATA SINGLE-POINT FOR O_2 DATA)
THERMOX WDG III CONTROL STATION
BOOSTER FAN MOTOR
INLET SAMPLE LINE
INLET
INLET
INLET
MANUAL TEST PORT
INLET

Figure 5. Layout of the CEM.

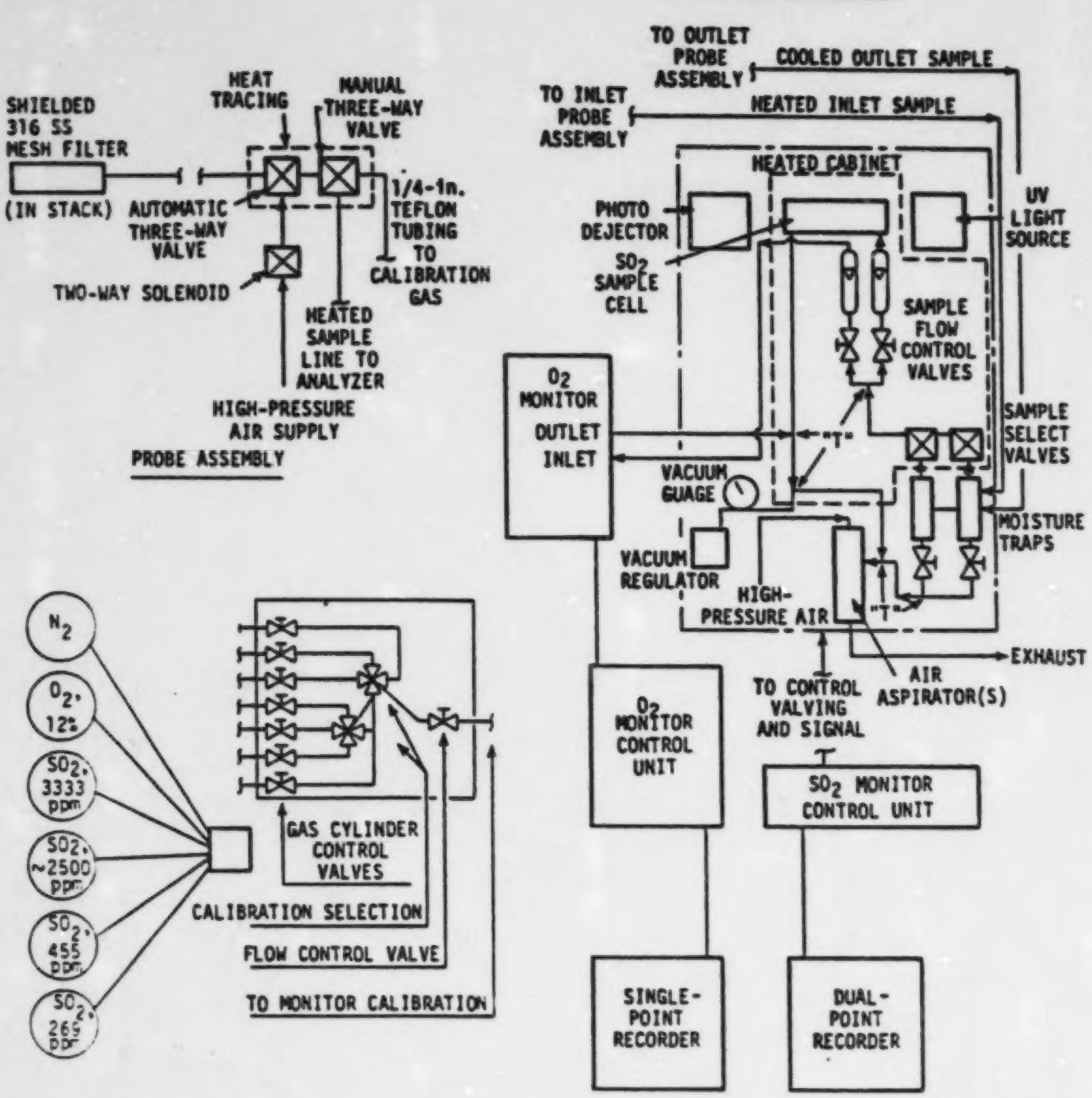
ANALYSIS SYTEM OF CEM
TO OUTLET PROBE ASSEMBLY
COOLED OUTLET SAMPLE
TO INLET PROBE ASSEMBLY
HEATED INLET SAMPLE
HEATED CABINET
PHOTO DEJECTOR
SO2 SAMPLE CELL
UV LIGHT SOURCE
SAMPLE FLOW CONTROL VALVES
SAMPLE SELECT VALVES
O2 MONITOR
OUTLET
INLET
"T"
VACUUM GUAGE
MOISTURE TRAPS
VACUUM REGULATOR
HIGH-PRESSURE AIR
"T"
EXHAUST
AIR ASPIRATOR(S)
TO CONTROL VALVING AND SIGNAL
O2 MONITOR CONTROL UNIT
SO2 MONITOR CONTROL UNIT
SINGLE-POINT RECORDER
DUAL-POINT RECORDER
SHIELDED 316 SS MESH FILTER
(IN STACK)
HEAT TRACING
MANUAL THREE-WAY VALVE
AUTOMATIC THREE-WAY VALVE
1/4-in. TEFLON TUBING TO CALIBRATION GAS
TWO-WAY SOLENOID
HEATED SAMPLE LINE TO ANALYZER
HIGH-PRESSURE AIR SUPPLY
PROBE ASSEMBLY
N2
O2, 12%
SO2, 3333 ppm
SO2, ~2500 ppm
SO2, 455 ppm
SO2, 269 ppm
GAS CYLINDER CONTROL VALVES
CALIBRATION SELECTION
FLOW CONTROL VALVE
TO MONITOR CALIBRATION

Figure 6. Simplified schematic of CEM.

Normally, one FGD monitoring system measures both FGD inlet and outlet gas streams on a 10-minute cycle. A sidestream sample for the oxygen monitor is taken parallel to the SO_2 analysis cell. Sample dewpoint is controlled by temperature-constant mist Knockout traps upstream from SO_2 and O_2 monitoring. Dewpoints are normally held to a maximum of approximately 37.8°C (100°F) and 17.5 Kilopascals (5 inches Hg) vacuum.

Monitoring instruments used have included DuPont 400 and 460 SO_2 monitors; Thermox WDG III, Beckman 742, and MSA 802 oxygen analyzers; and Leeds/Northrup speedomax and Esterline Angers MS-401-BB recorders. Mention of these instruments by name does not imply an EPA endorsement; rather, these products are mentioned factually as the particular models EPA contractors were using at the time of testing. The principles of operation of the SO_2 and O_2 analyzers are as follows:

Dupont 400/460 - measures SO_2 by ultraviolet spectrophotometry. This is accomplished by drawing a sample into a windowed cell, passing ultraviolet light through the sample, and measuring the photometric output obtained in a wavelength specific to SO_2 absorption.

Thermox WDG III - measures O_2 by electrochemical means across a heated zirconium oxide cell.

Beckman 742 - reduces oxygen between two electrodes causing a current flow. The magnitude of the current is proportional to tne partial pressure of O_2 present in the sample stream.

MSA Model 802 - operates on the principle that oxygen is paramagnetic while most other gases are dimagnetic. This effect on a magnetic field which develops with movement of oxygen in the sample stream is translated directly into the amount of oxygen present.

Equipment at each continuous monitoring site were certified according to guidelines specified by Performance Specifications 2 and 3 in the Federal Register, Vol. 44, No. 197, October 10, 1979. EPA Methods 3 and 6 were performed to complete relative accuracy tests. To certify monitors for continuous operation, the systems successfully passed tests for calibration error, drift, response time, relative accuracy, and a 168 hour conditioning period in which no modification or maintenance, except that specified as routine by the equipment manufacturers, was performed.

RESULTS OF MONITORING

Data acquired at two industrial sites--Parma and Rickenbacker--were examined to determine the relative success in obtaining continuous SO_2 data with FGD system and boiler reliability.[6,7] These results are tabulated in Table 6 along with utility data acquired in previous EPA monitoring programs.[8,9,10]

What is noteworthy from Table 7 is the superior FGD and SO_2 monitor reliability for industrial systems as compared to utility systems, the one exception being the prototype utility system at Shawnee. Industrial boiler FGD reliability ranged from 89 percent at Parma No. 3 to 95 percent at Parma No. 1 and Rickenbacker. This compares to 39 percent at Conesville "A", 68 percent at Pittsburgh, and 95 percent at Louisville "S".

Also worth mentioning are the monitoring systems reliabilities for industrial boilers, ranging from 98 percent at Parma 3 and 94 percent at Parma 1 to 80 percent at Rickenbacker. This shows that for sodium systems, the monitor may be as reliable as the FGD system, while monitors on lime based systems are considerably less reliable. For the lime systems at utility sites, monitoring reliability ranged from 85 percent at Shawnee to 37 percent on Conesville "B".

TABLE 7. BOILER, FGD, AND MONITOR PERFORMANCE

Boiler size	Total time of operation, hrs	Boiler downtime, hrs	FGD downtime, hrs[1]	Monitor downtime, hrs[1]
INDUSTRIAL				
Parma #1	456	0	22.75	22
Parma #3	1968	1345	67.25	13.25
Rickenbacker	1320	2	69	252
UTILITY				
Louisville N	7582	1444	414	3851
Louisville S	9414	1444	414	2019
Pittsburgh I & III	3890	1245[2]	1245[2]	2646
Conesville A	4416	528	2448	908
Conesville B	4416	528	2376	888
Shawnee TCA	1176	0	0	168

[1] Abnormal operation included as downtime.

[2] Boiler and FGD are considered one unit, since no bypass allowed.

It should be noted, however, that all tests were performed using existing monitoring systems where possible, and that location of probes, mist carryover, and age and repair status of monitors were all factors contributing to monitoring downtime. Many of these problems may be eliminated when installing a new FGD and monitoring system.

Tables 8 and 9 examine monitoring failures at Parma and Rickenbacker, respectively.[11-12] At Parma, 21 of 35 hours (21 of 25 data hours or over 80 percent of data loss) were lost because of a timer malfunction that controlled sample gas feed. At Rickenbacker, outlet probe plugging, a common ailment of calcium-based FGD monitoring systems, contributed to 45 percent (131 of 292 data hours) of all lost data. Faulty chart drive also contributed to a substantial data loss--55 hours or 18 percent of downtime.

DATA REPORTING

SO_2 emission data are collected and reported in a manner compatible with the requirements of 40 CFR 60, Subpart Da as outlined in the June 11, 1979, Federal Register. This procedure consists of converting analyzer outputs for sulfur dioxide and oxygen concentrations to mass emission rates using F-factors as follows:

$$E = \frac{CFK}{1-M} \times \frac{20.9}{\frac{20.9-O_2}{1-M}}$$

Where E = Emission factor - lb/million Btu

C = SO_2 concentration - ppmv, wet basis

TABLE 8. DATA LOSS AT PARMA

No. of hours lost	No. of occurrences	Cause	Remedy
2	2	Chart drive of O_2 recorder stuck	Reset paper winding
7	1	Chart paper out	Added chart paper
1	1	Air valve stuck	Replaced seals
21	2	Minor cycle timer stuck	Unstuck timer
4	1	Probe maintenance	Not applicable

TABLE 9. DATA LOSS AT RICKENBACKER

No. of hours lost	No. of occurrences	Description	Remedy
4	1	Testing of gas conditioner	N/A
17	1	Timer off	Turned timer on
3	1	Zero card maintenance	N/A
5	4	Integrity, audit, calibration checks	N/A
55	10	Chart drive sticking, O_2 recorder	Repaired chart
2	1	Change ultraviolet lamp bulb	N/A
16	4	Outlet interface failure and correction	N/A
131	13	Outlet probe filter plugged	Changed filter
1	1	SO_2 sample cell dirty	Cleaned cell
12	1	Oil in plant air	Plant performed maintenance
1	1	Inlet blowback repairs	N/A
23	1	Blowback valve leak	Repaired leak
7	1	Blowback solenoid valve failure	Replaced valve
10	1	Inlet probe plugged	Replaced probe
5	1	O_2 recorder chart paper out	New roll paper
TOTAL 292	42		

[a] Not applicable.

F = Stoichiometric conversion factor, 9820 dscf/million Btu for subbituminous coal

K = Conversion factor, 1.659×10^{-7} lb/dscf/per ppmv

O_2 = Oxygen concentration, percent by volume as measured

M = Moisture fraction as measured (for dried samples, M=0)

The sulfur dioxide and oxygen concentration results are obtained by multiplying the strip chart readings as a percent of scale by the appropriate calibration factor. Moisture content is assumed to be that for a saturated gas stream at the analyzer temperature and pressure.

The emission factor is calculated for each FGD system inlet and outlet test point. The sulfur dioxide removal efficiency for a module is calculated by:

$$\text{Efficiency} = \frac{\text{Ein} - \text{Eout}}{\text{Ein}} \times 100 \text{ percent}$$

Data Listings

15-Minute Readings--

Computer printouts list complete 15-minute readings showing the wet basis FGD inlet and outlet SO_2 and O_2 concentrations, the moisture content of each gas stream tested, and the corrected (dry basis) inlet and outlet SO_2 and O_2 values. Inlet and outlet emission rates (lb $SO_2/10^6$ Btu) and FGD system efficiencies are calculated and listed on an hourly basis.

1-hour Averages--

The 1-hour averages of inlet and outlet emission rates and FGD efficiency are calculated from 15-minute readings for any hour when data for at least two 15-minute periods are captured.

24-Hour Averages--

The 24-hour (daily) averages of inlet and outlet emission rates and FGD system efficiency are also calculated for data days that meet a minimum data capture requirement of 18 hours. Table 10 shows a typical 24-hour average data listing.[13]

Omitted Data

Normally there are three basic reasons for which data periods do not appear in listings or calculations:

Absence of Data Caused by Process Shutdown or CEM Failure--

During process shutdown or CEM failure, no data are available for listing. When only one parameter (inlet or outlet readings) is omitted, the other, if available, is listed. These instances are caused by failure of the sampling interface in the case of the affected parameter, but continued operation in the case of the unaffected parameter. The FGD efficiency data cannot be calculated in such cases.

Unrepresentative Process Operation--

When temporary loss of scrubber feed results in loss of SO_2 emission control, data are not included in the listings.

Failure to Obtain a Sufficient Data Base for Computation--

At least two 15-minute readings per hour are necessary for an hourly reading to be computed in the initial data listing, and at least eighteen 1-hour averages of a parameter are needed for data to be included in the summaries of 1-hour and 24-hour results.

TABLE 10. TYPICAL RESULTS BASED ON 24-HOUR SO_2 AVERAGES

```
      17 OPERATING DAYS
      SUMMARY OF RESULTS
      USING 24-HOUR AVERAGES

      LOCATION: GM PARMA BOILER 1
      DATE:  2-28-80

  DATE           LOAD    E         E         EFF
                          IN        OUT

   2-28-80        74.    3.808      .449     88.2
   2-29-80        80.    3.995      .383     90.3
   3- 1-80        73.    3.939      .379     90.4
   3- 2-80        74.    3.817      .273     92.8
   3- 3-80        76.    3.241      .226     93.0
   3- 4-80        68.    3.076      .313     89.9
   3- 5-80        67.    2.870      .218     92.4
   3- 6-80        68.    3.579      .320     90.8
   3- 7-80        66.    4.205      .236     94.6
   3- 8-80        56.    4.657      .310     93.0
   3- 9-80        53.    4.049      .370     90.8
   3-12-80        71.    3.891      .190     95.2
   3-13-80        67.    3.775      .401     89.8
   3-14-80        65.    4.004      .492     87.6
   3-15-80        54.    4.211      .311     92.6
   3-16-80        67.    3.616      .253     93.0
   3-17-80        65.    3.979      .315     92.1

 # AVERAGES       17.  17.        17.       17.
 % OF DATA        55.  55.        55.       55.
 MINIMUM          53.    2.870      .190   87.580
 MAXIMUM          80.    4.657      .492   95.192
 MEAN             67.    3.809      .320   91.535
 STD.DEV.          7.     .437      .084    2.107
 % STD.DEV.       11.   11.472    26.215    2.301

 30-DAY  REMOVAL EFICIENCY USING
      24
 (MEAN)    E   , E       : 91.587%
      30  IN    OUT

              Y
 NOTE: (MEAN)   IS DEFINED AS:
              X
      THE MEAN FOR X DAYS USING Y-HOUR AVERAGES
```

```
      25 OPERATING DAYS
      SUMMARY OF RESULTS
      USING 24-HOUR AVERAGES

      LOCATION: GM PARMA BOILER 3
      DATE:  2- 4-80

  DATE           LOAD    E         E         EFF
                          IN        OUT

   2- 5-80        32.    3.568      .144     95.9
   2-11-80        34.    2.862      .150     94.8
   2-12-80        34.    2.909      .183     93.7
   2-13-80        36.    2.895      .161     94.5
   2-14-80        39.    2.741      .187     93.2
   2-15-80        40.    2.963      .170     94.1
   2-16-80        40.    2.988      .087     97.1
   2-17-80        40.    2.824      .092     96.7
   3-18-80        41.    3.804     1.038     73.7
   3-19-80        37.    3.903      .796     79.2
   3-20-80        38.    4.399      .469     89.3
   3-21-80        41.    3.789      .195     94.9
   3-22-80        43.    3.830      .142     96.3
   3-23-80        37.    3.984      .162     95.9
   4- 1-80        46.    4.505      .356     92.2
   4- 2-80        42.    4.651      .412     91.1
   4- 3-80        40.    4.546      .359     92.1
   4-16-80        39.    5.315      .256     95.1
   4-17-80        39.    4.853      .321     93.2
   4-18-80        35.    3.824      .308     92.0
   4-21-80        32.    3.822      .323     91.6
   4-22-80        32.    3.976      .308     92.3
   4-23-80        31.    3.780      .251     93.4
   4-24-80        35.    3.634      .298     91.8
   4-25-80        32.    3.849      .348     91.1

 # AVERAGES       25.  25.        25.       25.
 % OF DATA        81.  81.        81.       81.
 MINIMUM          31.    2.741      .087   73.663
 MAXIMUM          46.    5.315     1.038   97.093
 MEAN             37.    3.769      .301   92.203
 STD.DEV.          4.     .695      .214    5.210
 % STD.DEV.       11.   18.453    71.193    5.650

 30-DAY  REMOVAL EFICIENCY USING
      24
 (MEAN)    E   , E       : 92.023%
      30  IN    OUT

              Y
 NOTE: (MEAN)   IS DEFINED AS:
              X
      THE MEAN FOR X DAYS USING Y-HOUR AVERAGES
```

DATA ANALYSIS AND RESULTS OF CONTINUOUS SO_2 MONITORING TESTS

The purpose of statistically analyzing SO_2 data obtained in this study is in response to the Clean Air Act Amendments of 1977, Section 111 which requires both a percent reduction in potential emissions and a fixed emission limit using the "best technological system of continuous emission reduction...". With the mandate of continuous monitoring of emissions where feasible, the problem then becomes one of determining: (1) the percentage reduction in emissions which can be consistently achieved by the best system of emission reduction and (2) an appropriate averaging time for reporting of emission reduction data. Of lesser importance for FGD systems is a statistical analysis of the actual emission rate since this rate is normally set to allow full utilization of high sulfur coal reserves and is seldom the controlling standard where FGD is selected as the best available control technology.

The purpose, therefore, of the following analyses is to provide the relationship between emission characteristics of properly designed and operated FGD systems and the frequency at which these systems will achieve given performance levels as a function of various averaging periods. It should be pointed out emphatically here that FGD system performance is not the only basis for SO_2 standards. In fact, the main reference cited herein is primarily devoted to analyzing coal sulfur variability, which becomes of paramount importance where it is determined that an FGD system cannot be economically or environmentally justified.

Previous analyses of utility FGD data have established that both normal and log normal distribution of emissions data may be used to statistically describe FGD performance.[14] For simplicity, the data examined in this paper are assumed to be normally distributed and will be described by four parameters:

° mean - a measure of the center of measurements, in this case, the arithmetic, long-run averages of measurements:

$$\overline{X} = \frac{\Sigma Xi}{N}$$

° standard deviation (SD) - a measure of the variability of measurements about the center, defined as the root-mean-square:

$$SD = \sum \frac{(Xi-\overline{X})^2}{N}$$

° relative standard deviation (RSD) - the ratio of the standard deviation to the mean, normally expressed as a percentage, e.g. $RSD = \frac{SD}{\overline{X}}$

° autocorrelation (AC) - the measure of a dependence between successive measurements. A correlation of +1 indicates perfect correlation, that is, successive measurements are identical. A correlation of zero indicates no dependence between successive measurements. Correlations below zero indicate that high measurements are followed by low and low by high. A high autocorrelation has the effect of increasing the likelihood of excursions above a threshold value for a given averaging period.

Summary statistics for inlet, outlet, and efficiency data are presented in Table 11 for one hour averages for the GM-Parma, Rickenbacker, and GM-St. Louis facilities.[15] The high variability at the Mead facility due to lack of operator control has been excluded here, as the results are not felt meaningful, although the Mead system did achieve a 96.1 geometric mean removal efficiency.

The results shown in Table 11 show that removal efficiencies were all above 90 percent, that the variability of emissions (as measured by the standard deviation) is reduced significantly by FGD (an average for the four systems of 61 percent), and that short-term data, one-hour averages, are highly dependent on the previous value. FGD systems also tend to reduce the autocorrelation (an average of 18% reduction for the four systems) of hourly averages somewhat.

TABLE 11. INDUSTRIAL FGD STATISTICS - ONE HOUR AVERAGES

Unit	N	Mean	SD	RSD(%)	AC
Parma 1 inlet	423	3.80	0.55	14.4	0.91
Parma 1 outlet	422	0.33	0.19	57.0	0.79
Parma 1 Eff. %	422	91.3	4.98	5.5	0.79
Parma 3 inlet	558	3.79	0.82	21.7	0.97
Parma 3 outlet	555	0.24	0.12	52.4	0.77
Parma 3 Eff. %	555	93.8	2.80	3.0	0.67
Rickenbacker inlet	1150	5.24	0.41	7.8	0.90
Rickenbacker outlet	1009	0.40	0.24	55.4	0.71
Rickenbacker Eff. %	971	91.9	4.58	5.0	0.70
St. Louis inlet	789	5.45	0.43	7.9	0.90
St. Louis outlet	790	0.19	0.20	70.8	0.74
St. Louis Eff. %	743	96.3	2.90	3.0	0.74

Table 12 shows the effects of averaging time on RSD.[16] Basically, RSD's decline as average time increases, and decline slower when autocorrelations are higher, as in the inlet RSD's as compared to outlet. Table 13 compares removal efficiency data for industrial boilers with utility data previously investigated, on a 24-hour basis.[17] Except for the adipic acid enhanced system (Shawnee venturi) and the very low sulfur coal application (Lawrence), utility FGD systems show significantly lower efficiencies than the industrial FGD systems examined. The majority of utility systems are lime based while 3 of 4 industrial systems are sodium based. The one industrial FGD system using lime, however, shows removal efficiency, variability, and autocorrelation characteristics similar to the sodium systems. One utility FGD system using a sodium (Wellman-Lord) absorbent exhibited low variability as did the adipic acid enhanced system, but the majority of utility systems exhibited instability when compared to the industrial systems.

In every instance, the utility systems were significantly auto correlated compared to the industrial systems; the one exception being Pittsburgh II for which the autocorrelation number is suspect (no consecutive 24-hr data were available for that system). This indicates that factors influencing FGD performance have a longer term effect in utility systems than for industrial boilers.

This study offers no explanation for the above phenomena, as it is likely due to combination of design and operating practices. The trends discussed herein are consistent with few exceptions.

The conclusion is that industrial lime, dual alkali, and sodium systems show inherently higher SO_2 removal, lower variability, are less susceptible to upsets, and are more amenable to monitoring than their lime-based

TABLE 12. EFFECT OF AVERAGING TIME ON RSD

Unit	Averaging time, hrs				
	1	24	360	720	1440
Parma 1 inlet	14.4	10.5	3.4	2.5	1.7
Parma 1 outlet	57.0	30.9	8.7	6.2	4.4
Parma 1 Eff. %	5.5	3.0	0.8	0.6	0.4
Parma 3 inlet	21.7	19.4	8.8	6.4	4.6
Parma 3 outlet	52.4	27.2	7.6	5.4	3.8
Parma 3 Eff. %	3.0	1.3	0.4	0.3	0.2
Rickenbacker inlet	7.8	5.5	1.8	1.3	0.9
Rickenbacker outlet	55.4	25.8	7.1	5.0	3.5
Rickenbacker Eff. %	5.0	2.3	0.6	0.4	0.3
St. Louis inlet	7.9	5.6	1.8	1.3	0.9
St. Louis outlet	70.8	34.7	9.6	6.8	4.8
St. Louis Eff. %	3.0	1.5	0.4	0.3	0.2

TABLE 13. COMPARISON OF REMOVAL EFFICIENCY DATA FOR UTILITY AND INDUSTRIAL BOILERS

Unit	Mean	SD	RSD	Auto-Correlation
UTILITIES				
Louisville North	83.8	4.7	5.6	.70
Louisville South	82.3	5.9	7.2	.69
Pittsburgh I	80.3	4.6	5.7	.47
Pittsburgh II	85.1	3.4	4.0	-.14
Chicago	89.1	1.3	1.5	.70
Shawnee TCA	88.3	2.2	2.5	.60
Shawnee Venturi	95.8	1.5	1.6	.89
Conesville A	84.7	6.1	7.2	.71
Conesville B	91.7	3.5	3.8	.63
Lawrence	93.6	5.3	5.7	.64
INDUSTRIAL BOILERS				
GM Parma 1	91.3	2.7	3.0	.11
GM Parma 3	93.8	1.2	1.3	.06
Rickenbacker	91.9	2.1	2.3	.07
GM St. Louis	96.3	1.4	1.5	.08

counterparts in the utility sector. These factors simplify considerably the development of an NSPS for industrial boilers where FGD systems may be justified environmentally and economically.

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APPLICABILITY OF FGD SYSTEMS TO OILFIELD STEAM GENERATORS AND SODIUM WASTE DISPOSAL OPTIONS

Presentation at EPA
Flue Gas Desulfurization Symposium
October 31, 1980
Houston, Texas

by

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ABSTRACT

This paper summarizes the evaluation of twelve commercial FGD processes for their applicability to steam generators at thermally enhanced oil recovery (TEOR) sites in California. All the FGD processes were compared at a common design basis with the sodium-throwaway process currently used at the TEOR sites. PEDCo concluded that the ammonia, lime, limestone, double alkali, and Chiyoda T-121 processes are very competitive with the sodium-throwaway process. The Department of Energy (DOE) sponsored the study in order to assist the oil companies in FGD process selection, and DOE in planning future R&D projects. The paper also presents the status of existing FGD systems in the oilfields.

The primary concerns about the sodium-throwaway process are the cost of the alkali and waste disposal. The paper reviews the current practices of sodium-based waste disposal in industrial applications which include disposal to holding pond for evaporation, and wastewater treatment and discharge to city sewer, tailing pond, or waste wells. The paper discusses two major options for bleed stream treatment: oxidation (oxidation/crystallization, or oxidation/partial quenching of the flue gas), and central regeneration (chemical or thermal). The marketability of Na_2SO_4, gypsum, or sulfuric acid will determine the applicability for a specific site.

APPLICABILITY OF FGD SYSTEMS TO OILFIELD STEAM GENERATORS AND SODIUM WASTE DISPOSAL OPTIONS

INTRODUCTION

The purpose of this paper is to summarize currently available flue gas desulfurization (FGD) processes and to evaluate their applicability to steam generators at thermally enhanced oil recovery (TEOR) sites in California. The paper also discusses some disposal techniques for the waste stream generated by sodium-throwaway FGD systems.

The California Air Resources Board (CARB) has imposed stringent regulations on the sulfur dioxide (SO_x) emitted from steam generators at the TEOR sites in Kern County, California. In anticipation of the regulations, the U.S. Department of Energy (DOE) sponsored this study to help the oil companies select an FGD process and to assist DOE in planning future research and development (R&D) projects.

PEDCo assessed twelve FGD processes: sodium carbonate, ammonia, limestone (conventional), limestone (Chiyoda T-121), lime, double alkali, dry scrubbing, citrate, sodium sulfite (Wellman Lord), magnesium oxide, carbon absorption (BF/FW), and copper oxide sorption (Shell/UOP). The first seven are nonregenerable and last five are regenerable. A regenerable process is defined here as one that both regenerates reagent and produces a salable byproduct. The evaluation showed that the regenerable FGD processes were too expensive and complex to be feasible at the TEOR sites. Therefore, only the nonregenerable processes are discussed in this paper. All the processes were compared with the sodium-throwaway process that is being widely used in the oilfields.

We first give an overview of FGD activity at California TEOR sites. We then discuss the evaluation approach, present FGD system cost estimatess and highlight the recommendations of the DOE study. Finally, we discuss current and other possible disposal practices for the effluent from the sodium-throwaway process.

CALIFORNIA OILFIELDS[1]

This section summarizes TEOR operations in California, the CARB SO_2 regulations, and FGD systems now in place in the oilfields.

TEOR Operations

According to a 1977 California Department of Conservation report, the San Joaquin Valley Air Basin accounted for about 80 percent of all TEOR operations in California. Of these operations, 99 percent are in Kern County, within a 30-mile radius of Bakersfield.

In a TEOR operation, steam is produced in small steam generators and injected into the oil wells to force the oil to flow as an emulsion with the condensate. Feedwater may be recycled to the generators from the crude oil recovery area, where the oil and water are separated, or water may be pumped from separate freshwater wells. Some of the oil produced is burned in the generators; usually about 3 barrels of oil is recovered by the steam produced by burning 1 barrel of crude. Although the composition and heating values vary, the fuel oil has a typical sulfur content of 1.2 percent and a heating value of 150,000 Btu/gal.

The oilfield operations in Kern County are of two basic types. Small, independent companies operate a few generators isolated from each other in the field, and big companies operate several steam generators in groups of 6 to 10. The generators are of two sizes: 20×10^6 Btu/h heat output, and 50×10^6 Btu/h heat output.

Several oil companies are involved in the TEOR operations, and in the past few years SO_2 emissions from the steam generators have increased significantly. The CARB estimates that by the end of 1979, TEOR operations in Kern County were burning about 70,000 bbl/day of oil and generating SO_2 emissions of about 250 tons/day.

Sulfur Dioxide Regulations

The CARB regulations of SO_2 emissions, which became effective in October 1979, apply to steam generators in Kern County with heat inputs of at least 15 million Btu/h.

The regulations specify that new sources (those with construction permits dated on or after February 21, 1979) may emit no more than 0.06 lb sulfur/million Btu input (about 0.12 lb SO_2/million Btu input). By July 1, 1982, existing sources (those with construction permits dated before February 21, 1979) may emit no more than 0.25 lb sulfur/million Btu input (about 0.50 lb SO_2/million Btu input). For a new steam generator producing typical crude oil (with a heating value of 150,000 Btu/gal and sulfur content of 1.15 percent), the regulations require about 90 percent SO_2 removal.

Kern County's current regulations require disposal of liquid and solid wastes in sites specified as Class 2-I. Such sites are permitted to contain nonhazardous wastes. The California Department of Health has undertaken a study to analyze wastes produced by sodium-throwaway scrubbers. The results of the study may cause the liquid waste to be classified as hazardous. In such a case, the material must be disposed of in sites specified as Class I.

In addition to the SO_2 regulations, CARB has also promulgated stringent NO_x emission regulations for the existing and new steam generators. These regulations are not, however, discussed in this paper.

Flue Gas Desulfurization

In October 1979, 79 sodium-based FGD systems were operating on 183 steam generators with an overall capacity of 9×10^9 Btu/h. By 1982, about 210 systems are expected to operate on nearly 620 generators with an overall capacity of 28×10^9 Btu/h. Table 1 summarizes the FGD systems at TEOR sites in California. All but one of the systems now in operation are using a sodium hydroxide or sodium carbonate (soda ash) scrubbing process.

The companies range from a small, independent operator using one scrubber on one generator to a company like Getty Oil using 11 scrubbers on a total of 88 generators. The scrubbers range from a manually controlled, simple eductor design to a fully instrumented, tray tower absorber with a complete array of auxiliary equipment. Most of the companies have purchased scrubber systems from system suppliers; two large companies, Getty and Mobil, have chosen to keep their FGD projects totally in-house, from system design to equipment installation and startup.

Currently operating systems are consistently removing at least 90 percent of the SO_2 in the generator flue gas. Across the industry, FGD reliability has been high, much higher on average than that of FGD systems in utility

TABLE 1. SUMMARY OF OPERATING AND PLANNED FLUE GAS DESULFURIZATION SYSTEMS IN THE CALIFORNIA OILFIELDS[a]

Status	No. of systems	No. of steam generators	Total capacity, 10^6 Btu/h output
Operating	79	183	9,000
Definitely planned	61	174	8,200
Early planning[b]	70	263	10,900
Total[c]	210	620	28,000

[a] As of October 1979.

[b] Several systems are not reported by oil companies because the plans are not definite.

[c] Approximate numbers.

power plant. One reason for this generally successful record is the use of sodium-throwaway processes for SO_2 removal in TEOR scrubbers. In contrast, over 90 percent of the U.S. utility FGD capacity utilizes lime/limestone scrubbing. Another reason is the size of the operation: the average TEOR FGD system controls flue gas equivalent to that from a 15 MW unit, while the average utility FGD system controls emissions from a boiler generating 400 to 500 MW. The simplicity of operation associated with the much smaller size enhances the FGD reliability.

EVALUATION APPROACH

PEDCo followed the following steps in evaluating each process:

Review of available literature: An information base was established from in-house files and published literature; additional information needed to complete the task was obtained by contacting knowledgeable sources. Flow diagrams of the FGD process were prepared for oilfield applications.

Examination of development and application: Whether the process has been demonstrated on a steam generator with an output of 50 million Btu/h or more was examined, along with any special requirements.

Calculation of material balances: The flow rate and composition of each process stream were defined for the common design basis shown in Table 2. Individual and overall material balances were calculated on the basis of key design requirements,, such as amount of SO_2 removed, reagent requirement, and bleed-off requirements.

Sizing of equipment items and determination of energy requirements: Individual material balances and design conditions were used to size process equipment items and determine operating energy requirements. Table 3 lists the design assumptions (such as type of absorber) for each of the nonregenerable processes.

Estimation of capital investment and annual cost requirements: For meaningful economic comparisons, a cost basis suitable for the TEOR sites was established.

Identification of design considerations: Critical design aspects of the process were identified and important operating variables were noted.

Assessment of environmental considerations: Waste disposal requirements were examined and problems pertinent to TEOR site application were assessed.

TABLE 2. DESIGN BASIS FOR A PROPOSED FGD SYSTEM
IN A CALIFORNIA OILFIELD APPLICATION

Steam generator characteristics	
Duty	300 million Btu/h output, total six generator bank (each @ 50 million Btu/h output), 380 million Btu/h input
Steam rate	290,000 lb/h, 80 percent quality
Excess air	15 percent
Load factor	90 percent (7884 h/yr)
Operating life	20 years
Fuel oil properties	
Sulfur, wt percent	1.14
H.H.V.	150,000 Btu/gal
Flow rate	20,040 lb/h for six generators
Flue gas characteristics @ FGD system inlet	
Total flow rate	133,800 acfm @ 500°F
Molecular weight	29.05 (wet basis), 30.44 (dry basis)
CO_2, vol percent	12.2
H_2O, vol percent	11.2
O_2, vol percent	2.6
Sulfur dioxide	600 ppmv (440 lb/h)
Emission regulation	
Maximum 0.06 lb sulfur/million Btu input ($\cong$0.12 lb SO_2/million Btu input)	
Degree of pollutant removal	
SO_2	90 percent
Ash	Not specified
NO_x	Nil
Miscellaneous	
Reheat	Stack gas reheat is not provided
Bypass	Bypass ductwork is not provided
Redundancy	Only slurry handling pumps are spared (100 percent)

TABLE 3. DESIGN ASSUMPTIONS FOR NONREGENERABLE FGD PROCESSES IN A CALIFORNIA OILFIELD APPLICATION

FGD process	Type	SO_2 absorber			Byproduct/waste management
		Gas velocity, ft/s	Pressure drop, in. H_2O	Liquid-to-gas ratio, gal/10^3 acf	
Sodium carbonate	Tray tower	9.0	7.0	6	Disposal in lined pond
Ammonia	Packed horizontal unit	11.4	7.0	8[a]	Disposal in lined pond
Limestone (conventional)	Spray tower	9.0	6.0	40	Disposal in lined pond
Limestone (Chiyoda T-121)	Jet bubbling reactor	b	b	b	Gypsum production[c]
Lime	Spray horizontal unit	18.2	3.0	20[a]	Disposal in lined pond
Double alkali	Tray tower	9.0	7.0	6	Disposal in lined pond
Dry scrubbing	Spray dryer	2.2	10.0	0.42	Haulage of waste[d]

[a] Value for each stage.

[b] Proprietary information.

[c] An interim pond with 2-year life provided for storage of gypsum.

[d] It is assumed that the supplier of trona would haul the waste back for disposal in the mines.

COST ESTIMATES[1]

This subsection presents estimated capital and annual costs. All costs are given in mid-1979 dollars. When necessary, costs have been escalated according to the average annual cost indexes taken from "Chemical Engineering."

Capital Investment

Total capital investment is the sum of fixed capital investment (FCI) and a contingency allowance of 15 percent of the FCI. The fixed capital investment consists of total direct investment (TDI) and total indirect investment (TII). The TDI includes total purchase equipment cost (PEC) and other direct investment costs.

The PEC's are obtained from various sources after sizing each equipment item. The other direct investment costs cover structural work, foundation, piping, electrical work, instrumentation, insulation and painting, and site preparation. Each of these is assumed to be a percentage of total PEC of an FGD system; the percentage is based on in-house data on FGD systems. The costs of a lined pond (including land for the pond) and vessel lining (including installation) are calculated separately for each system because they do not involve other direct costs. These costs are added to the total PEC and other direct investment costs to obtain the TDI.

The TII for an FGD system includes engineering costs, contractor's fee, interest during construction, field overhead, freight and other offsides, insurance and taxes, and spares. Each indirect investment cost is expressed as a percentage of TDI.

Annual Costs

Total annual costs of an FGD system consist of direct and indirect operating costs. The direct operating cost covers cost of raw material, utilities, operating labor, supervision, maintenance and repairs, and sludge/waste disposal. The indirect operating cost consists of overheads, fixed costs, and general expenses.

Results

Table 4 summarizes the capital investments and annual costs of nonregenerable processes.

TABLE 4. SUMMARY OF CAPITAL INVESTMENTS AND ANNUAL COSTS[a,b]
(mid-1979 dollars)

	Capital investment		Net annual costs	
FGD process	Total, $ × 10^3	Unit, $/scfm	Total, $ × 10^3/yr	Unit, $/bbl oil burned
Sodium carbonate	966.6	13.2	591.7	1.24
Ammonia	1000.5	13.6	530.6	1.12
Limestone (conventional)	1469.1	19.9	580.8	1.22
Limestone (Chiyoda T-121)	1379.4	18.7	571.4	1.20
Lime	1522.8	20.7	593.8	1.25
Double alkali	1528.9	20.8	614.2	1.29
Dry scrubbing	2378.5	32.3	826.9	1.74

[a] The accuracy of estimates ranges from +30 to -10 percent.

[b] Design basis: Six 50 x 10^6 Btu/h generators manifolded into one FGD system, 1.14 percent sulfur oil, 90 percent SO_2 removal, no reheat, no bypass, only slurry handling pumps spared.

Total Capital Investment--

The sodium carbonate process has the least capital investment, although ammonia scrubbing is only slightly more expensive. All other processes, except dry scrubbing, require about 40 percent more capital investment. The capital investment for dry scrubbing is significantly higher than for other processes because of the fabric filter it includes.

Annual Costs--

Ammonia scrubbing is the least expensive process on an annual basis, whereas all other processes (except dry scrubbing) are very competitive with the sodium carbonate process. Again, the annual costs for dry scrubbing are the highest because of its high capital investment.

RECOMMENDATIONS[1]

The ammonia scrubbing process has the lowest annual cost even when the ammoniacal liquor, which can be sold as a fertilizer byproduct, is disposed of in a lined pond. The mildly acidic liquor may be especially suitable for the alkaline soil in the San Joaquin Valley. The primary disadvantage of ammonia scrubbing is the potential for blue haze formation. PEDCo recommends that a pilot plant be operated in the oilfields to evaluate the feasibility of fumeless ammonia scrubbing. In view of the NO_x regulations imposed by CARB, the pilot plant could be integrated with a selective catalytic (or noncatalytic) reduction unit upstream of the wet scrubber.

Lime, limestone, Chiyoda T-121, and double alkali processes are very competitive with the sodium-throwaway process, and PEDCo recommends that the oil companies consider them. Chiyoda T-121 has the advantage of producing gypsum, which is used by the farmers in the valley.

As oil prices increase, the oil companies may find it cost-effective to burn coal in the steam generators. Low-sulfur coal can be transported to California from Wyoming and Utah. In this case, dry scrubbing will be preferable to all other processes because it also offers very high particulate removal efficiency. The application of fluidized bed combustion (FBC) may also be evaluated .

Finally, there is a growing need to evaluate ways to treat the waste stream from the sodium-throwaway systems.

WASTE STREAM TREATMENT

More than half the FGD systems on industrial boilers in the United States use the sodium-throwaway process. Although these installations have demonstrated high reliability and high SO_2 removal efficiency, absorbent costs and possible restrictions on current waste disposal practices may reduce the applicability of the process. In this subsection, the current disposal practices are reviewed and other possible disposal techniques are discussed.

Current Waste Disposal Methods[2]

The bleed stream from a sodium-throwaway FGD system contains both sodium sulfite and sodium sulfate, and may contain inerts or ash. Also, if taken before addition of the alkali makeup, the bleed stream contains sodium bisulfite. All the sodium salts are highly water soluble, and the bisulfite and sulfite have a chemical oxygen demand (COD). The effluent flow from industrial boiler FGD systems has to meet state or local water quality standards. Although these standards vary from state to state, they would typically require control of pH, suspended solids, and COD.

Most of the larger industrial FGD systems send the bleed stream through the wastewater treatment facility on site. The treatment involves clarification to remove fly ash and solids, adjustments of pH, and aeration to oxidize sulfite species to sulfate for COD reduction. The solution is discharged to municipal sewers, rivers, or tailings pond.

The most common disposal technique is discharge of untreated bleed stream to a lined or unlined evaporation pond. All of the FGD systems using this method are in the Western states where the annual evaporation rate exceeds the annual rainfall. Finally, some of the FGD systems in the California oilfields dispose of the waste liquor by injection in the exhausted deep wells.

Bleed Stream Treatment Techniques

The bleed stream from a sodium-throwaway FGD system may be processed on site to produce sodium sulfate or in a central facility to produce gypsum or sulfuric acid. These techniques are discussed below.

Sodium Sulfate Production--

The bleed stream is sent through a clarifier to remove solids, and oxidized by air to obtain a solution of sodium sulfate. It may be necessary to adjust the pH of the bleed stream before the oxidizer. This solution is sent through a chiller/crystallizer and a dryer to produce sodium sulfate; a part of the clear liquor is disposed of and the other part recycled to the FGD system. This technique, shown in Figure 1, is similar to the purge stream treatment section of the Wellman-Lord process. Vapor compression, reverse osmosis, or the combination of these two unit operations may also be used for producing sodium sulfate.

Alternatively, the solution of sodium sulfate can be contacted with the hot flue gas in a quencher upstream of the FGD system. This variation is shown schematically in Figure 2. Because the bleed stream flow rate is usually not sufficient to allow saturation of the flue gas, the gas is partially quenched and sodium sulfate is formed as a powder. If the quencher is designed as a spray dryer, most of the powder is collected at the bottom of the quencher. Sodium sulfate particulates that remain in the flue gas can be removed by a venturi at the inlet of the absorber. Two critical areas of design would be maintaining the temperature of the partially quenched flue gas above its acid dew point, and keeping the particle size of the powder coarse enough to avoid excessive pressure drop in the venturi.

The advantages of either technique include elimination of liquid waste disposal problems and reduction of the water requirement of the FGD system. The disadvantages include the increase in capital and operating costs due to the added equipment and the need to sell sodium sulfate. Therefore, the marketability of salt cake should be examined at each site when investigating economic feasibility.

Central Regeneration--

In an area where several industrial boilers use caustic scrubbing, a central regeneration facility can be installed to receive the bleed streams from the FGD systems. The spent liquor is clarified and reacted with lime slurry to regenerate the caustic and produce a calcium sulfite/sulfate slurry. Such a treatment is identical to the regeneration step of a double alkali process. The slurry is thickened and the caustic returned to the individual FGD system. The calcium sulfite/sulfate sludge is oxidized in the presence of sulfuric acid and dewatered to produce gypsum. This technique is shown schematically in Figure 3.

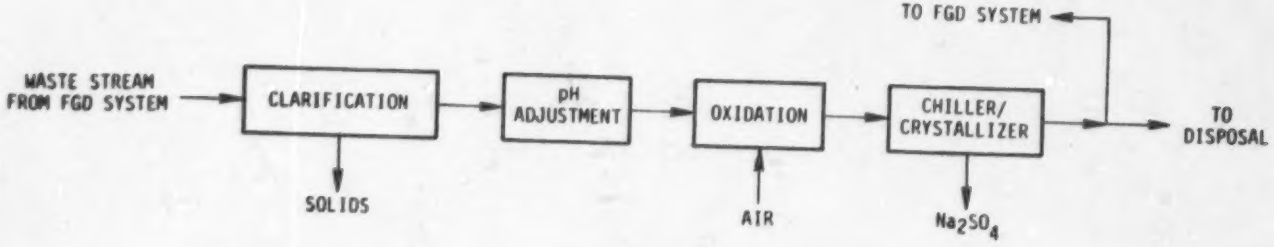
WASTE STREAM FROM FGD SYSTEM
CLARIFICATION
SOLIDS
pH ADJUSTMENT
OXIDATION
AIR
CHILLER/ CRYSTALLIZER
Na_2SO_4
TO FGD SYSTEM
TO DISPOSAL

Figure 1. Onsite sodium sulfate production by crystallization.

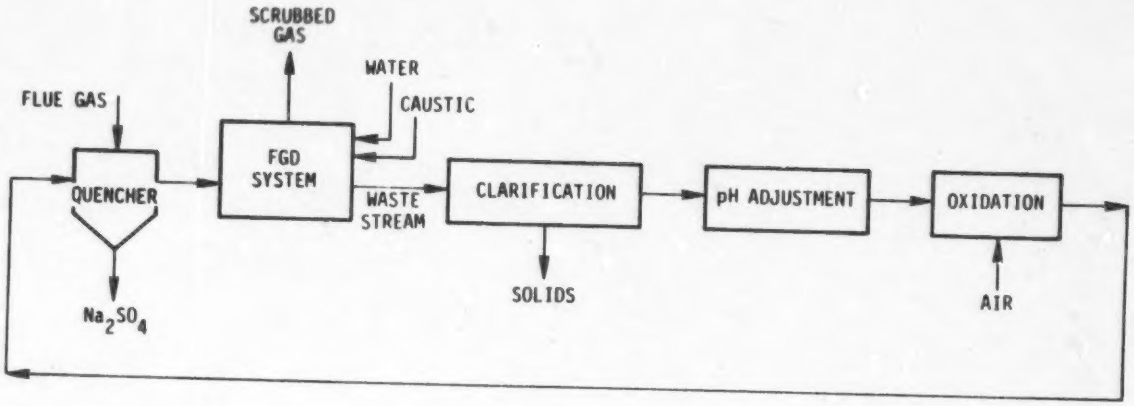
FLUE GAS
QUENCHER
Na_2SO_4
SCRUBBED GAS
FGD SYSTEM
WATER
CAUSTIC
WASTE STREAM
CLARIFICATION
SOLIDS
pH ADJUSTMENT
OXIDATION
AIR

Figure 2. Onsite sodium sulfate production by spray drying.

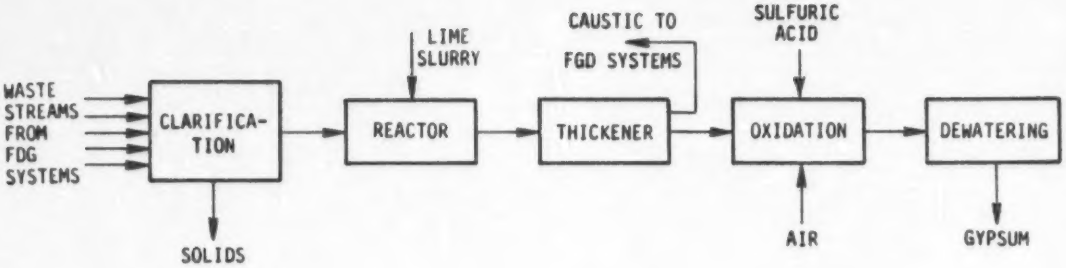
WASTE STREAMS FROM FDG SYSTEMS
CLARIFICATION
SOLIDS
LIME SLURRY
REACTOR
CAUSTIC TO FGD SYSTEMS
THICKENER
SULFURIC ACID
OXIDATION
AIR
DEWATERING
GYPSUM

Figure 3. Central chemical regeneration - gypsum production.

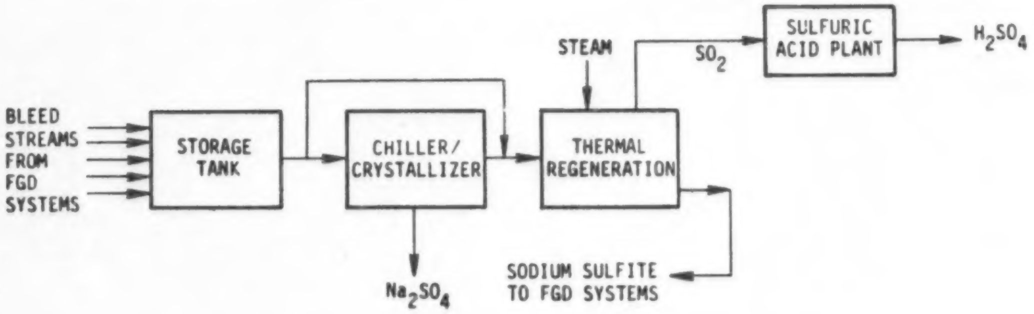
BLEED STREAMS FROM FGD SYSTEMS
STORAGE TANK
CHILLER/ CRYSTALLIZER
Na_2SO_4
STEAM
THERMAL REGENERATION
SO_2
SULFURIC ACID PLANT
H_2SO_4
SODIUM SULFITE TO FGD SYSTEMS

Figure 4. Central thermal regeneration - sulfuric acid production.

In an area where several new sodium-based FGD systems are planned for small industrial boilers, each system can use sodium sulfite scrubbing of the sort used in the absorption loop of a Wellman-Lord system. Each FGD system would have a prescrubber to remove ash and chlorides and a staged tray tower to maximize the sodium bisulfite/sulfite ratio in the bleed stream. Bleed streams from the individual systems can be thermally regenerated in a central facility shown schematically in Figure 4. Such a facility would produce sulfuric acid and sodium sulfate, both of which need to be sold. Also, the facility would produce sodium sulfite liquor for return to the individual FGD systems.

The economic feasibility of a central facility, presumably operated cooperatively by the several companies or by an independent company, depends on the marketability of the products, transportation costs for liquid materials, and assurance of a long-term commitment by the companies involved.

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PERFORMANCE EVALUATION OF AN INDUSTRIAL SPRAY DRYER FOR SO_2 CONTROL

by

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ABSTRACT

Using methods specified by EPA for compliance testing, the performance of a dry SO_2 control system was evaluated. The system tested treated flue gas from the coal-fired boiler located at the Amcelle Plant of the Celanese Fibers Company in Cumberland, Maryland. A brief description of this system and its operation is presented.

Results, based on 25 days of data obtained over a 33-day period, showed the mean SO_2 removal to be 69 percent, with a range of 60 to 80 percent, except for upsets, over the test period when the sulfur content of the coal averaged 2 percent. Operating experience with the spray-dryer-baghouse system is summarized for the 5-month period ending with the completion of testing on September 30, 1980.

PERFORMANCE EVALUATION OF AN INDUSTRIAL SPRAY DRYER FOR SO_2 CONTROL

INTRODUCTION

TRW, Inc., under contract to the Environmental Protection Agency, performed testing of the dry SO_2 control system serving the coal-fired (No. 5) boiler at the Amcelle Plant of the Celanese Fibers Company in Cumberland, Maryland. Installation and certification of instrumentation at the site for the performance testing were performed according to provisions for SO_2 compliance testing and began in late April 1980. The test program was concluded in September 1980.

This paper describes the results of this testing and discusses factors impacting testing and system operation. Since the spray-dryer/fabric-filter system for the flue gas cleaning (FGC) system tested is among the first such commercial units to become operational in the United States, particular attention is given to SO_2 removal performance and system operation and availability.

SYSTEM DESCRIPTION[1,2]

Celanese Fibers Company installed a coal-fired boiler in 1979 to supplement the existing oil- and gas-fired boilers at their Amcelle Plant in Cumberland, Maryland. This installation was undertaken to improve the economics of supplying process steam for the production of synthetic fiber. A spray dryer and fabric filter combination was chosen to provide flue gas desulfurization on the basis of cost, the lack of available space for ponding wastes from a wet FGD scrubber, and the need to provide good particulate control.

The flue gas desulfurization system was purchased as a turnkey installation from Rockwell International and Wheelabrator-Frye, Inc. A flow diagram of the system is presented in Figure 1.

Coal-Fired Boiler

The coal-fired water-tube boiler at the Amcelle Plant is identified as the plant's No. 5 boiler. The boiler is an Erie City spreader-stoker with a traveling grate for continuous ash discharge. This boiler had previously been retired from service at a Celanese plant in Rome, Georgia. The boiler was retubed when it was reconstructed at the Cumberland, Maryland (Amcelle) plant (Table 1).

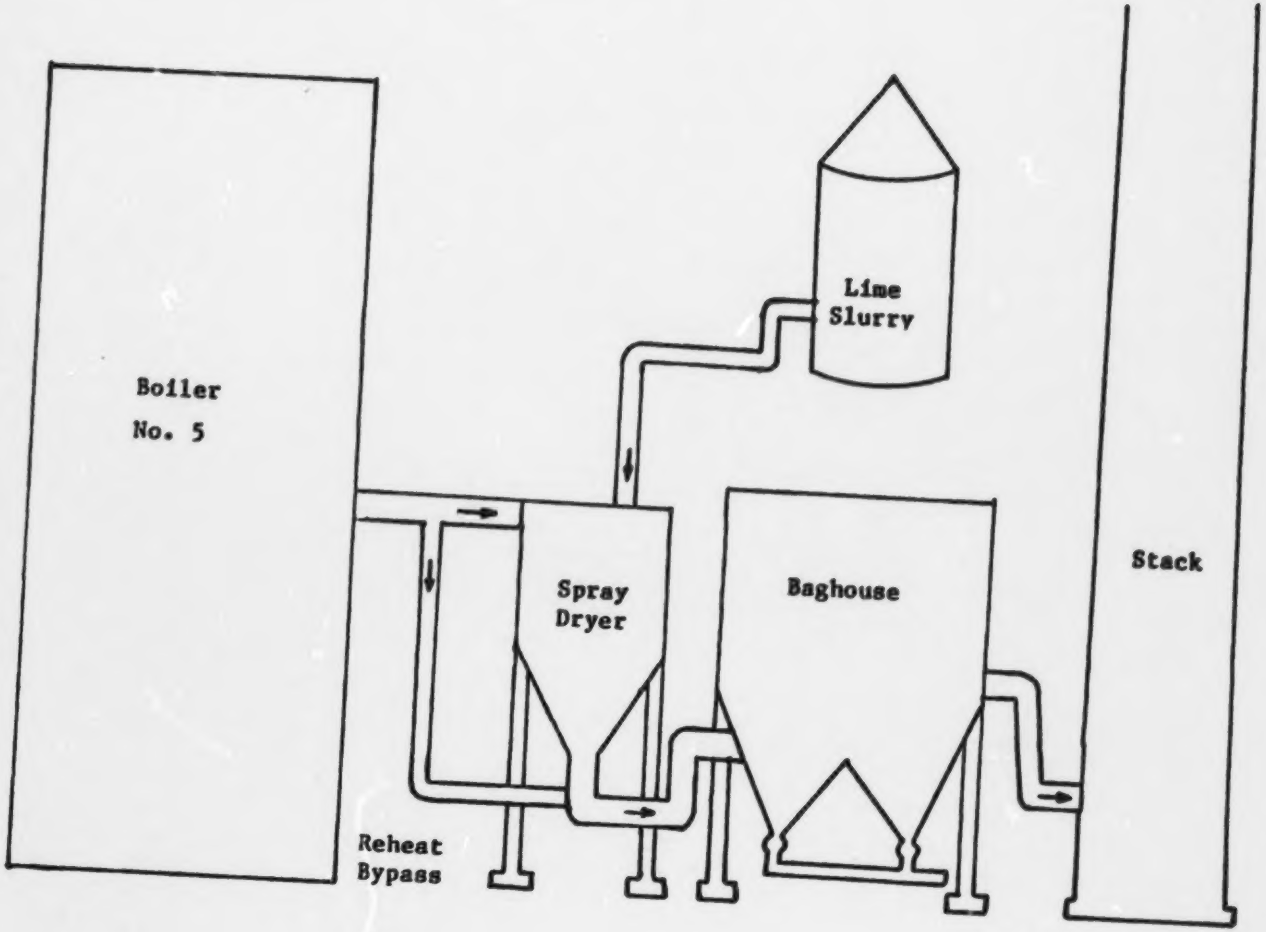
Boiler
No. 5
Lime
Slurry
Spray
Dryer
Baghouse
Stack
Reheat
Bypass

Figure 1. Flue gas desulfurization system.

Table 1. Boiler Data

Amcelle Plant Boiler No. 5

Boiler Type	- Erie City Spreader Stoker	
Fuel	- Coal and Natural Gas	
	Coal	Gas
Type	Bituminous	Natural Gas
Fuel Heating Value	29,056 kJ/kg (12,500 Btu/lb)	37.2 MJ/m^3 (1,000 Btu/ft^3)
Sulfur Content	1.0 to 2.0 percent	0.0 percent
Ash Content	8.0 to 20.0 percent	0.0 percent

The coal boiler is rated at 156 million kJ/hr (148 million Btu/hr) with secondary boiler fuels of gas/No. 6 fuel oil. At the boiler's maximum rating of 68,000 kg steam/hr (150,00 lb/hr) when fired by a combination of coal and oil or gas, the flue gas to be treated by the dry FGD system is 41.1 m^3/s (87,000 acfm) at 216°C (420°F). At the boiler's nominal coal-fired rating of 49,900 kg steam/hr (110,000 lb/hr), the flue gas to be treated is 30.7 m^3/s (65,000 acfm) at 193°C (380°F).

Analyses of randomly selected coal samples are presented in Table 2. The sulfur content of the coals received during the test period ranged between 1.25 and 2.76 percent, with a mean of 2.02 percent (dry basis).

Table 2. Selected Coal Analyses

Sample No.	Vol %	Ash %	Sulfur %	HHV kJ/kg	HHV Btu/lb
1 (8-22-80)	19.9	14.95	1.58	29,860	12,846
2 (8-29-80)	31.6	18.96	1.92	27,998	12,045
3 (9-12-80)	17.2	13.97	1.36	30,153	12,972
4 (9-23-80)	33.14	16.81	2.24	29,411	12,653

Table 3 illustrates flue gas design conditions for various coal firings.

Table 3. Flue Gas Characteristics

Amcelle Plant Boiler No. 5

Fuel Steam Production	Coal 49,900 kg/hr (110,000 lb/hr)
Gas Temperature	193°C (380°F)
Gas Flow Rate	30.7 m^3/s (65,000 acfm)
SO_2 Concentration	800 to 2,500 ppm
SO_2 Exhaust Rate	113 to 363 kg/hr (250 to 800 lb/hr)
Particulate Loading	8.5 to 11.9 g/m^3 (3.7 to 5.2 gr/dscf)[a]

[a] dscf denotes dry standard cubic feet.

Spray Dryer

The gas cleaning system is designed to provide flue gas desulfurization removals ranging from 70 percent for 1 percent sulfur coals to 87 percent for 2 percent sulfur coals from half to full boiler load. Most of this SO_2 removal takes place in the spray dryer where the SO_2-laden flue gas is passed through a finely dispersed fog of lime slurry and water.

The spray dryer consists of a single, 6.1-m (20-ft) diameter vessel containing a rotary atomizer (Figure 2). This rotary atomizer, or Bowen wheel, is driven at approximately 16,000 rpm. The lime slurry is fed to the wheel at a liquid-to-gas ratio of 0.04 l/m^3 (0.3 gal/1000 acf), where it is centrifugally dispersed into the gas stream. A swirling motion is imparted to the flue gas as it enters the top of the spray dryer through a fixed-vane rotary ring to increase turbulent mixing of the flue gas and the lime slurry.

Approximately 20 percent of the flue gas bypasses the spray dryer, thus providing reheat to raise the gas temperature prior to its entry into the fabric filter. This is necessary for dry operation and compensates for temperature drop in the fabric filter. The total quantity of water fed to the spray dryer is automatically adjusted to hold the gas temperature from the spray dryer at a set value.

Lime System

The lime system is depicted in Figure 3. The dry storage silo provides approximately a 10-day lime supply. High-calcium pebble quicklime is gravity fed into the lime slaker where it is mixed with water to provide a 20 to 30 percent (by weight) slurry. The lime system is designed to provide 125 percent of required capacity when the boiler is fired at its maximum rate with 2 percent sulfur coal and overfired gas/oil. Design features of the lime system include automatic flushing of the pumps and piping with water to prevent deposits.

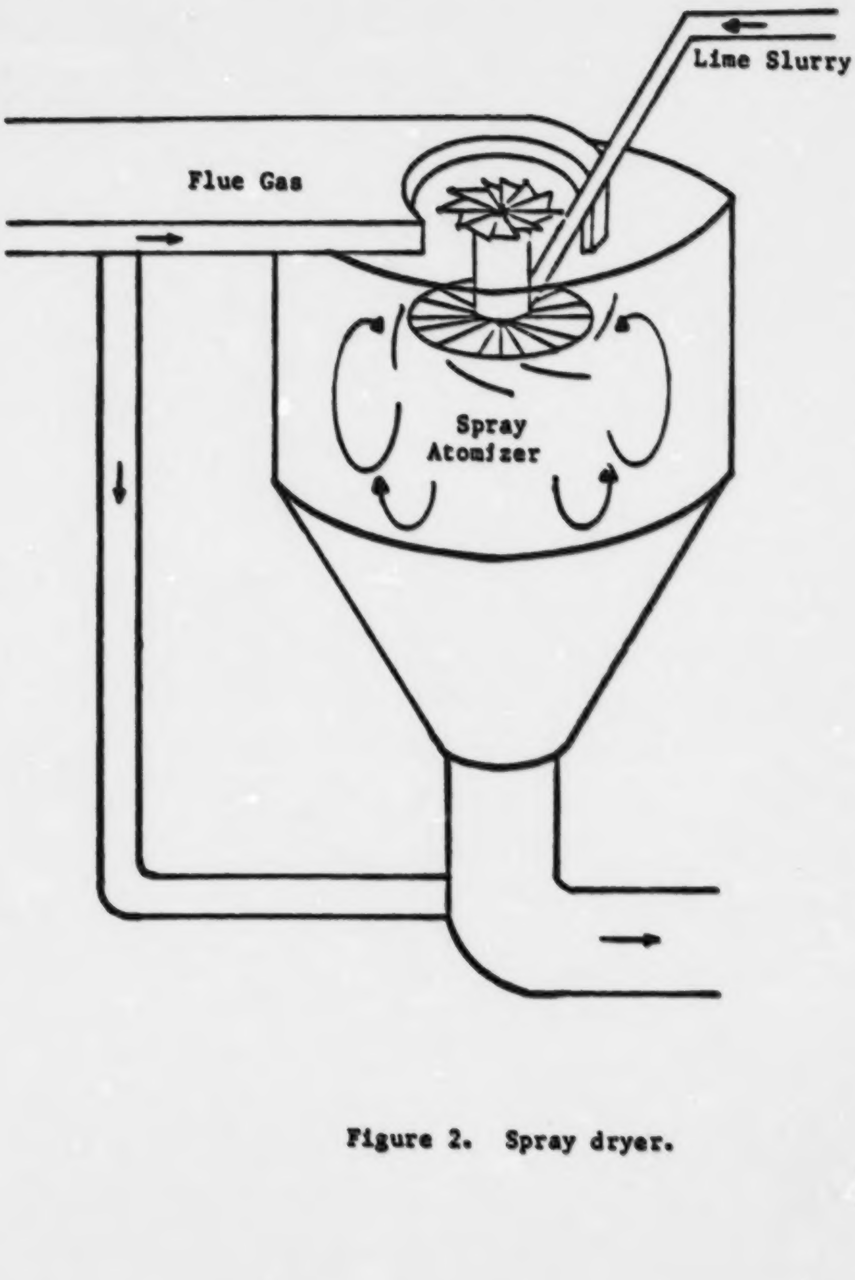
Lime Slurry
Flue Gas
Spray
Atomizer

Figure 2. Spray dryer.

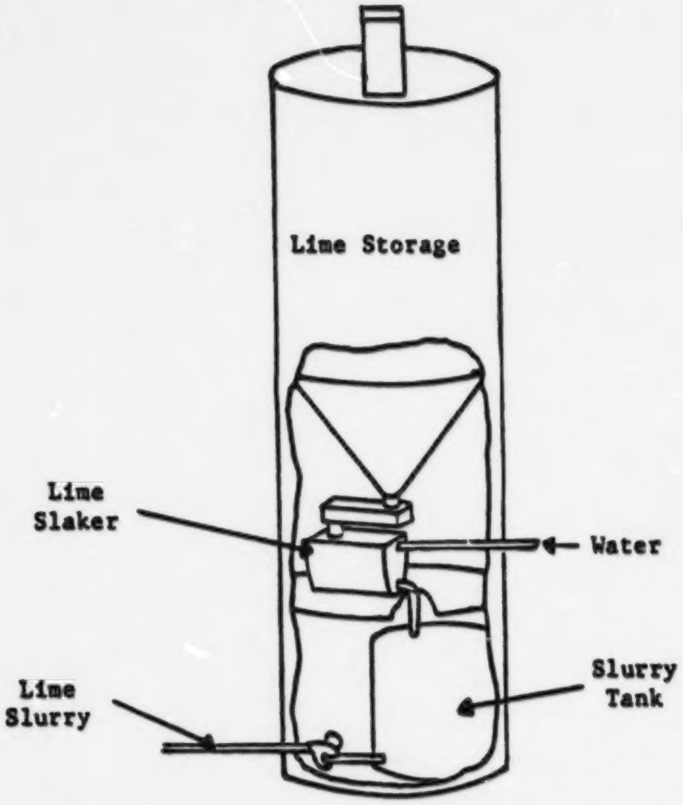
Lime Storage
Lime Slaker
Water
Slurry Tank
Lime Slurry

Figure 3. Lime system.

Fabric Filter

The fabric filter consists of a four-compartment pulse-jet baghouse manufactured by Wheelabrator-Frye, Inc., and is shown in Figure 4. Each compartment contains 225 bags. The baghouse is designed to operate with three compartments on-line when the boiler is operating at its nominal coal-firing rate to produce 49,896 kg/hr (110,000 lb/hr) of steam.

The air-to-cloth ratio will vary between 2.2 and 6.8 with a design pressure drop of 317.5 mm (12.5 in.) H_2O. The filter medium is a fiberglass-reinforced felt manufactured by Huyck.

OPERATING EXPERIENCE

Milestones in System Development

Celanese placed the order for the flue gas cleaning system in January 1979.[2] Construction of the system by Rockwell and Wheelabrator-Frye was completed in October 1979. Boiler installation was not completed, however, until mid-December 1979. Acceptance testing of the FGD system was completed on February 21, 1980.

TRW began collecting data for the demonstration test phase in May 1980. The objective of the program was to collect 30 days of continuous monitoring data, representing proper operation of the flue gas cleaning system, using compliance test methods. Problems with the boiler, the FGD system, and the continuous monitoring delayed completion of this test phase until September 30, 1980. This section describes the types of problems encountered in operating the dry FGD system and briefly summarizes the availability of the system.

System Availability

Preparation for the continuous monitoring program at the plant site began in April 1980. The boiler went down for refractory repairs on April 20. From that date through the end of the program on September 30, the boiler was off-line approximately 12 percent of the time. It was on-line but running abnormally an additional 5 percent. Thus, boiler problems prevented representative characterization of the FGD system for about 17 percent of the time TRW was on-site. This amounted to 672 hours out of 3,912 hours in the period.

The FGD system was off-line (not operating at all) about 23 percent of the time. The FGD system operated abnormally an additional 12 percent of the time. During this time the slurry feed rates were so low or unsteady that monitoring of any significant SO_2 scrubbing was prevented. Thus, the FGD system was unavailable 35 percent of the time, or a total of 1,354 hours.

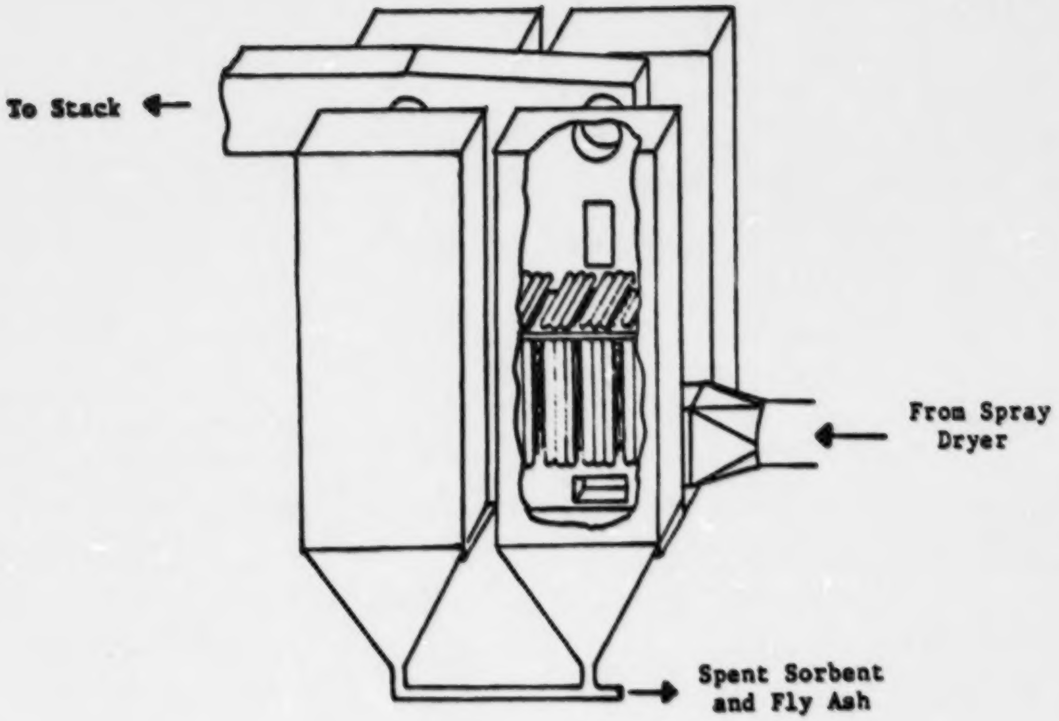
To Stack
From Spray Dryer
Spent Sorbent and Fly Ash

Figure 4. Fabric filter.

The availability of the system was significantly improved in September when most of the continuous monitoring data were collected. During this period the FGD system was off-line less than 19 percent of the time and operated abnormally an additional 8 percent of the time, giving an availability of 73 percent. Table 4 summarizes the availability of the boiler and FGD system during the test program.

The types of problems encountered and measures to resolve them by Rockwell International/Wheelabrator-Frye and Celanese are discussed below.

Types of Problems Operating problems encountered and their effects on the program were broken down among four system components. These were the boiler, the lime feed system, the spray dryer, and the fabric filter. Problems with each of these components impact the entire FGD system.

Steam Boiler The boiler operated entirely with eastern coal at about half of its rated load with relatively steady steam production for most of the monitoring period. After initial delays caused by necessary refractory repairs, the boiler went off-line once more during the test period for refractory repairs. This problem had no effect on the FGD system other than interrupting the source of flue gas. Another boiler problem which could have a serious impact on the FGD system is a boiler tube leak. A tube leak was suspected when one of the ducts in the FGD system became plugged due to moisture. When the plugging problem occurred, the boiler was taken off-line for inspection. No tube leaks were found so attention was turned to other potential sources of excess moisture in the flue gas.

Another problem which impacted the performance of the FGD system was the variability of coal quality. Coal sulfur content varied widely throughout the early part of the program. The quality became less variable near the end of the program, but proximate analyses of daily coal deliveries gave sulfur contents ranging from 1.25 to 2.76 percent. When operating in an automatic control mode to keep the outlet SO_2 concentration at a set value, the FGD system could respond to rapid changes in inlet SO_2 concentration within an hour so that hourly averages of emissions remained constant. Under manual control, the outlet SO_2 concentration followed the inlet concentration in the absence of operator adjustment. With uniform coal quality and automatic control of slurry flow, large fluctuations in inlet SO_2 concentrations were absent and a steady outlet SO_2 concentration was maintained.

Another problem which relates to coal supply involves the amount of fines in the coal. Coal fines, when suddenly dumped into the furnace, cause rapid changes in boiler load and the flue gas flow, changes in SO_2 emissions, and increased particulate matter and opacity levels in the stack. Fast changes in flue gas flow and SO_2 concentration make it difficult for the spray dryer to keep SO_2 emissions at a desired level. Such large and rapid load fluctuations occurred on September 5, 6, and 7, and the resultant effects on SO_2 emissions are documented in a later section.

Table 4. System Availability

Component	Availability*, %		
	Apr-Sep	Aug-Sep (720 hr)	Sep 25-30 (144 hr)
Boiler	82.8 (3240/3912)	93.3	100.0
FGD System	62.4 (2246/3600)	73.2	96.2
Spray Dryer	81.8	97.8	100.0
Lime Feed System	83.2	76.5	96.2
Fabric Filter	99.8	98.9	100.0

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*Availability is defined as percentage of time in the period that a component is operating normally.

One unpredictable problem with coal supply which affected SO_2 removal efficiency was a fire in the main coal silo. Because of this fire, boiler load had to be reduced temporarily as the silo was emptied. Another coal, higher in sulfur, was used until this problem was solved, resulting in a lower boiler load (18,144 kg/hr (40,000 lb/hr) of steam) for several days.

Lime Feed System The lime feed system uses a paste-type slaker with a grit-screen removal system. It was designed to supply sufficient slurry for SO_2 removal from coal up to 2 percent sulfur. The slurry is fed to the atomizer by progressing cavity pumps; under the design conditions, one pump is in use and one is a spare. However, to cope with the higher sulfur coals encountered, the single pump had to be operated at high speeds and this led to rapid pump wear. To alleviate this, the system was modified to use both pumps in parallel. This was the normal mode of operation throughout the latter part of the test period.

Most other problems with the lime feed system related to plugging somewhere in the system because of grit in the lime. Although grit was supposed to have been removed by screens inside the slaker, damaging quantities of it passed through or bypassed the screens into the rest of the system. Failure to remove grit caused plugging within the slaker, inside the flow lines, in the slurry pump, and in valves at various times in addition to excessive wear in the pumps. Dual element screen filters were eventually installed in the feed system, but not enough time elapsed before the end of the program to assess whether they solved the problem.

Spray Dryer There were three types of problems that occurred within the spray dryer. One of them resulted in the wetting of the dryer wall and discharge of damp material from the dryer. It was caused by maldistribution of lime slurry in the atomizer and was corrected by redesign.

The other two problems encountered with the spray dryer also related to the atomizer. The rotary atomizer was subject to clogging with grit particles if they were not screened sufficiently from the slurry. The spray dryer was shut down for cleaning when this clogging occurred. The other problem was failure of the bearings supporting the shaft of the atomizer wheel, caused by unbalance due to grit plugging the atomizer wheel.

Fabric Filter Some difficulties which affected performance of the FGD system were experienced in the fabric filter (baghouse). The most serious problem was the unexpectedly high pressure drop through the fabric filter. This was apparently caused by moisture on the bags which occurred during an upset and combined with ash and lime to form a coating that increased the resistance to flow. To lower the pressure drop through the baghouse, design and process changes were made, including increasing the pulse-jet air volume by approximately 15 percent. Tests since this modification indicate that this has solved the problem. The only other baghouse problem encountered related to periods when the compartments were opened for inspection or maintenance. Opening the baghouse compartments to do this work increases the air leakage into the baghouse which then alters the excess oxygen levels recorded and thus the corrected SO_2 concentrations. No effect on SO_2 removal was measured during bag replacement. There was, however, a reduction in SO_2 removal efficiency immediately after bag replacement, presumably due to an absence of filter cake containing lime on the new bags as opposed to the old ones.

Maintenance and Operating Needs

The two-stage dry FGD system as installed at the Celanese Fibers Company required 28 to 46 hours of maintenance each week and close operating supervision for continuous operation. Boiler operators handle the FGD system operation along with their other duties. Modifications made to the system after operating experience had been gained have the potential to make this a much more reliable system. As described above, most of the operating problems relate to plugging caused by grit in the slurry and water vapor condensing in the flue gas due to low operating temperatures. Both of these problems can be solved through changes in operation and design. Maintenance needs will also be reduced by these modifications.

Because of problems experienced thus far, however, redundancy of critical components is recommended. Specifically, three slurry pumps are needed with two on-line at all times and one as a spare. A spare atomizer is necessary to limit spray dryer shutdowns due to atomizer failure. Filters should be set up to provide uninterrupted slurry flow to the spray dryer during periods when one filter element is being replaced or cleaned. A means of keeping the outlet SO_2 monitor operating continuously is needed. This redundancy will permit a steadier outlet SO_2 level and more consistent FGD system performance via operation in the automatic control mode.

PERFORMANCE EVALUATION

Data on SO_2 removal, which were typical of fully operating dry FGD system performance and whose relative accuracy was fully documented, were collected only during the final month of the test program. This period of "good" data collection ran from August 28 through September 30, 1980. The boiler generally ran at a steady load (about half of the rated value because of seasonal reduction in steam requirements) throughout most of this period, and the FGD system operated almost continuously. Analysis of the system's SO_2 removal performance will focus on this period as the most representative portion of the entire program.

All inlet and outlet SO_2 monitors were installed and run according to 40 CFR 60, Appendix B. The analyzers were certified according to "Performance Specifications 2 and 3 for Continuous Monitors in Stationary Sources." This included a 168-hour conditioning period, a 168-hour operational testing period, system response tests, relative accuracy tests, and daily calibrations. All comparisons of continuous monitoring data with measurements made by EPA reference methods were within the required limits. Hourly averages of SO_2 emissions in parts per million (ppm) were calculated from a minimum of two data points per hour. These hourly averages were then corrected to zero percent oxygen dilution and converted to a "pounds per million Btu" basis. Calculations of SO_2 removal efficiency were then based on these average hourly SO_2 emission values.

Determination of SO_2 emissions in pounds per million Btu was done using the F-factor technique. An F-factor for dry flue gas from coal of 9820 dscf/10^6 Btu (263.9 m^3/GJ) was used. Heat input to the boiler was calculated from available data. Hourly averages of steam flow were used to derive hourly values of coal feed rate from daily totals of coal consumption. Figure 5 shows steam production to follow coal consumption throughout the monitoring period. The excellent correspondence of these variables supports the validity of using one of these parameters to determine the other.

Typical data for inlet and outlet SO_2 concentrations are shown in Figures 6 and 7. These data are representative of the 25 days when continuous monitoring methods met EPA's compliance criteria. These criteria include collection of data for over 18 hours daily with the FGD system treating boiler flue gas. Note that Figure 6 shows the outlet SO_2 concentration, which was measured in the stack, to follow closely the SO_2 concentration at the inlet to the spray dryer, and this curve indicates no corrective action being taken to adjust slurry flow rate for varying inlet SO_2 concentration. With the FGD system in the automatic control mode, the outlet SO_2 concentration (see Figure 7) was relatively constant, indicating that the slurry flow was adjusted to accommodate even rapid changes in inlet SO_2 concentration.

Although the FGD system was designed to operate automatically, this was not always possible because of frequent malfunctions in the stack SO_2 monitor which provided feedback to the spray dryer control system. Problems with this monitor necessitated frequent periods of manual operation. Under manual operation, the slurry flow sometimes became so high that the outlet concentrations were 50 ppm or less. On these occasions, SO_2 removal efficiencies exceeded 90 percent.

The average daily SO_2 removal efficiencies for the continuous monitoring period cited earlier are given in Figure 8. Except for periods of system upset, the removal efficiency ranged between 60 and 80 percent. These upsets resulted from problems discussed earlier. The only prolonged

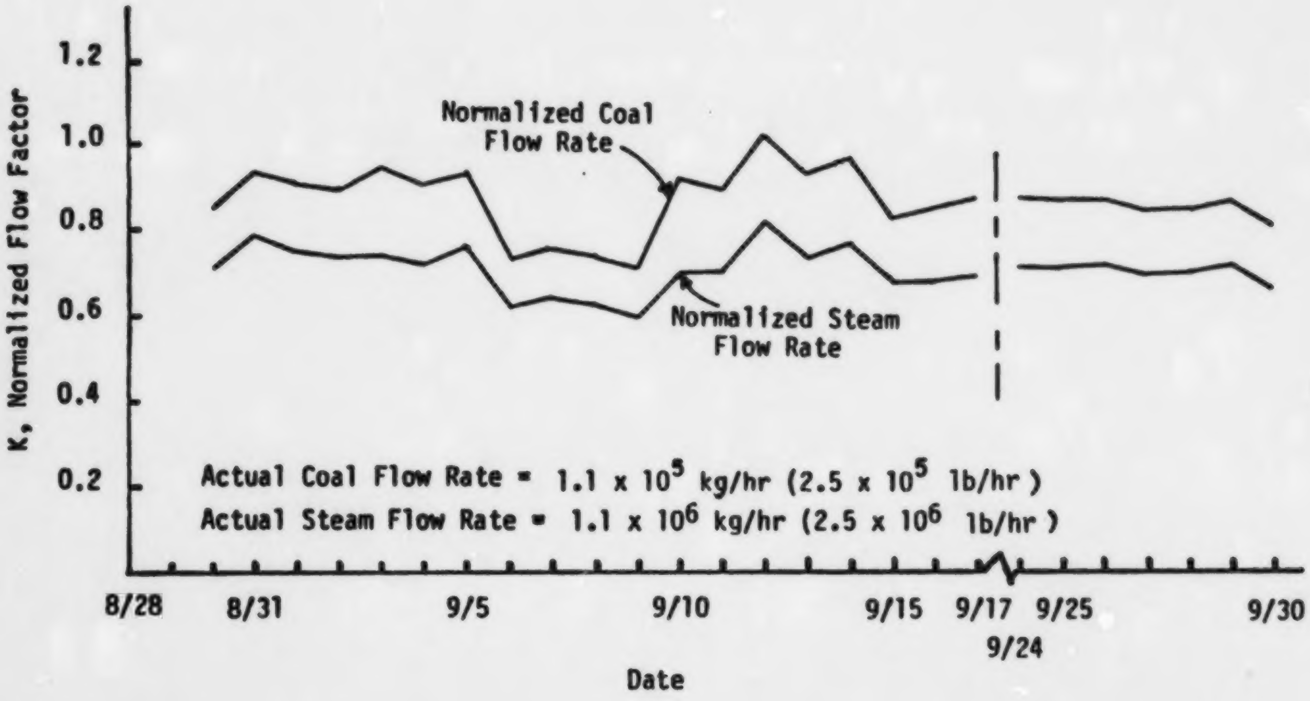

Figure 5. Relationship of steam production and coal consumption.

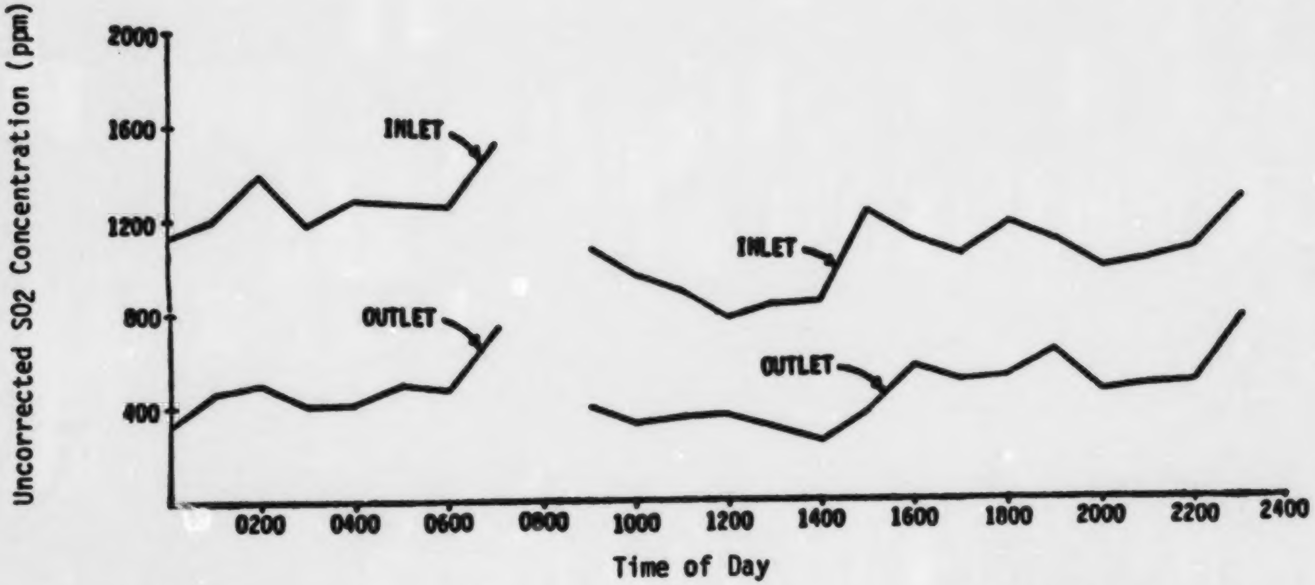

Figure 6. Average hourly SO_2 concentrations for September 3, 1980 with FGD system controlled manually.

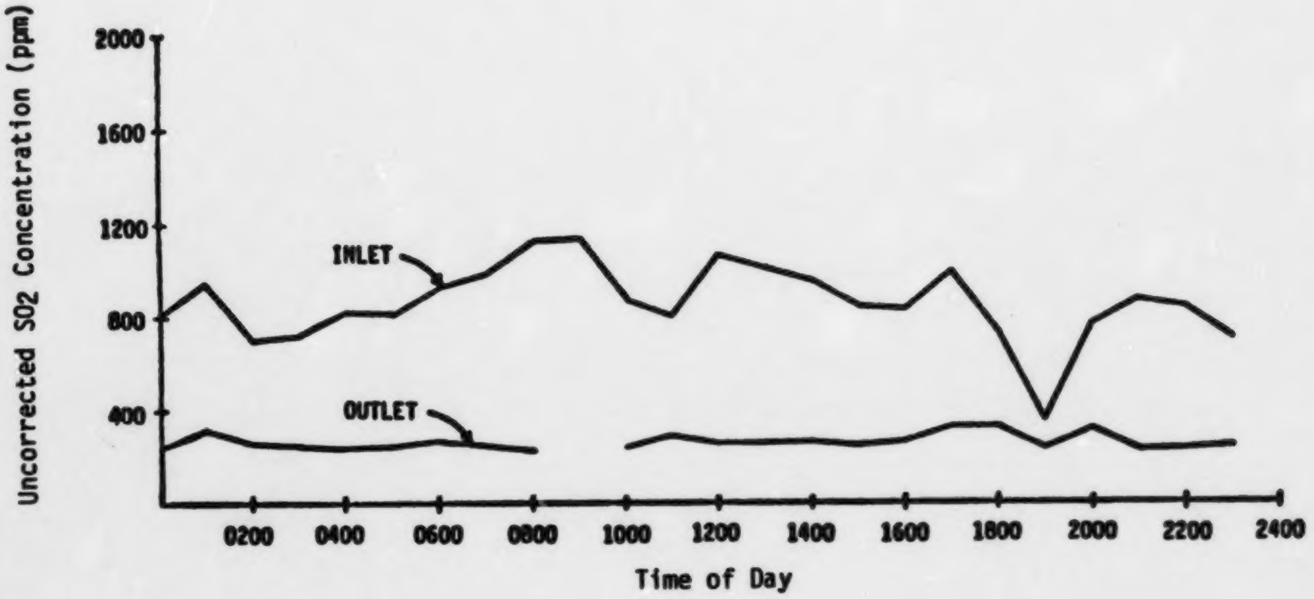

Figure 7. Average hourly SO_2 concentrations for September 8, 1980 with dry FGD system controlled automatically.

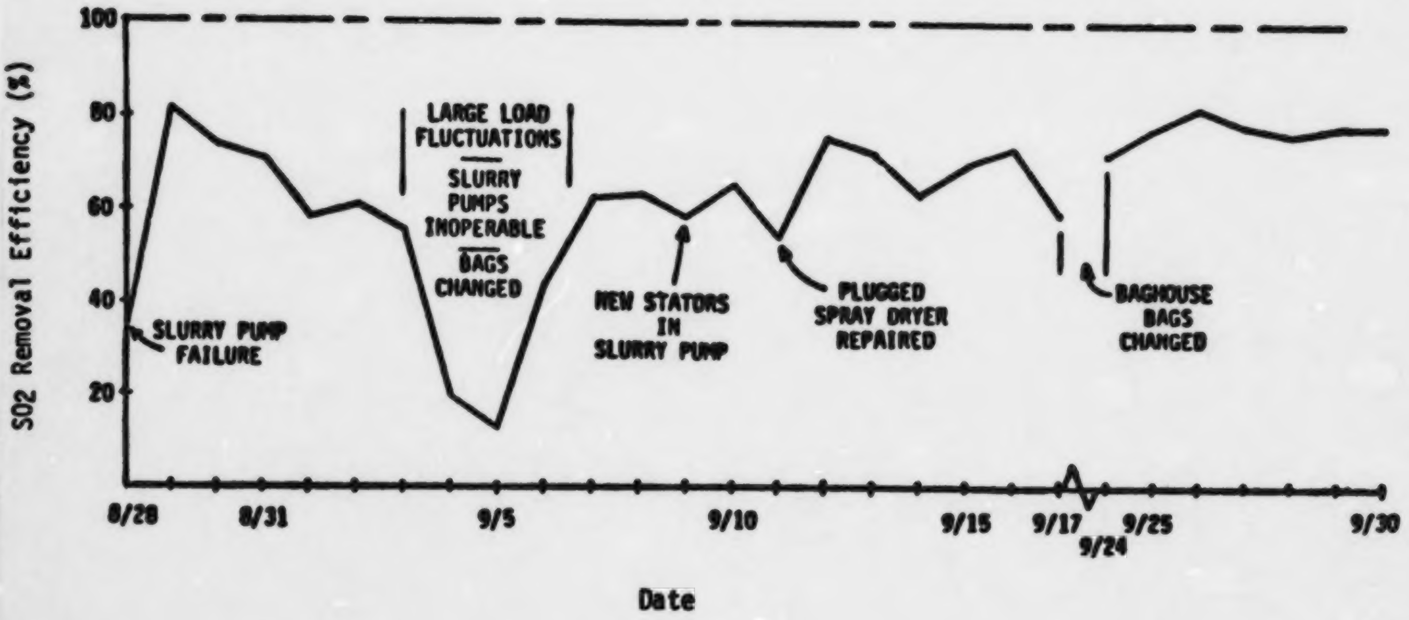

Figure 8. Average daily SO_2 removal efficiency for dry FGD system.

period of low SO_2 removal occurred between September 3 and 6 and stemmed from the inability to maintain steady boiler load in conjunction with slurry pumping problems. The mean SO_2 removal efficiency for the 25 days of performance data was 69 percent (see Figure 9), and the standard deviation of this mean was 9%. However, over nearly the last week of the test program (which was terminated prematurely because of failures of continuous monitoring equipment that precluded reasonably quick resolution), the average daily SO_2 removal efficiency remained near 80 percent, based on 23 hours of hourly averaged data for each day.

No parameteric tests were conducted during this program; hence, no quantitative correlations between operating variables (boiler heat input, flue gas flow rate, inlet and outlet SO_2 concentration, coal sulfur content and stoichiometric ratio) were determined. SO_2 removal was more a function of the spray dryer condition and performance than the boiler variables. In general, when the spray dryer with its lime slurry system was operating well, the SO_2 removal efficiency was satisfactory and independent of the boiler variables.

SUMMARY OF PREVIOUS TESTS AND REPORTED COSTS

A series of compliance and performance tests were completed in late February 1980 and showed the system to be in full compliance with federal, state, and local requirements.[2] The February tests confirmed that the flue gas cleaning (FGC) system guarantees requiring less than 37.5 kg/hr (70 lb/hr) of SO_2 emissions with the system automatically controlled, particulate matter emissions below 0.023 g/m^3 (0.01 grains/acf), and an overall FGC system pressure drop below 266.7 mm (10.5 in.) H_2O were being met.

Actual operating cost savings in switching from an oil- or gas-fired boiler to a coal-fired boiler met estimates, and daily lime costs equalled approximately $175.[2] During questioning following the presentation of their paper, Crowe et al.[2] also reported that cost savings using the coal boiler approximated several thousand dollars daily and that the stoichiometric ratio used was the design value.

Capital costs for the system are reported as $1.25 million,[3] but this sum does not include all system costs. Celanese provided the foundations, ash handling, and field wiring, while the given sum was for the spray dryer, baghouse, and lime system. Since the vendor and the buyer consider total capital and operating and maintenance costs to be proprietary, these costs have not been released.

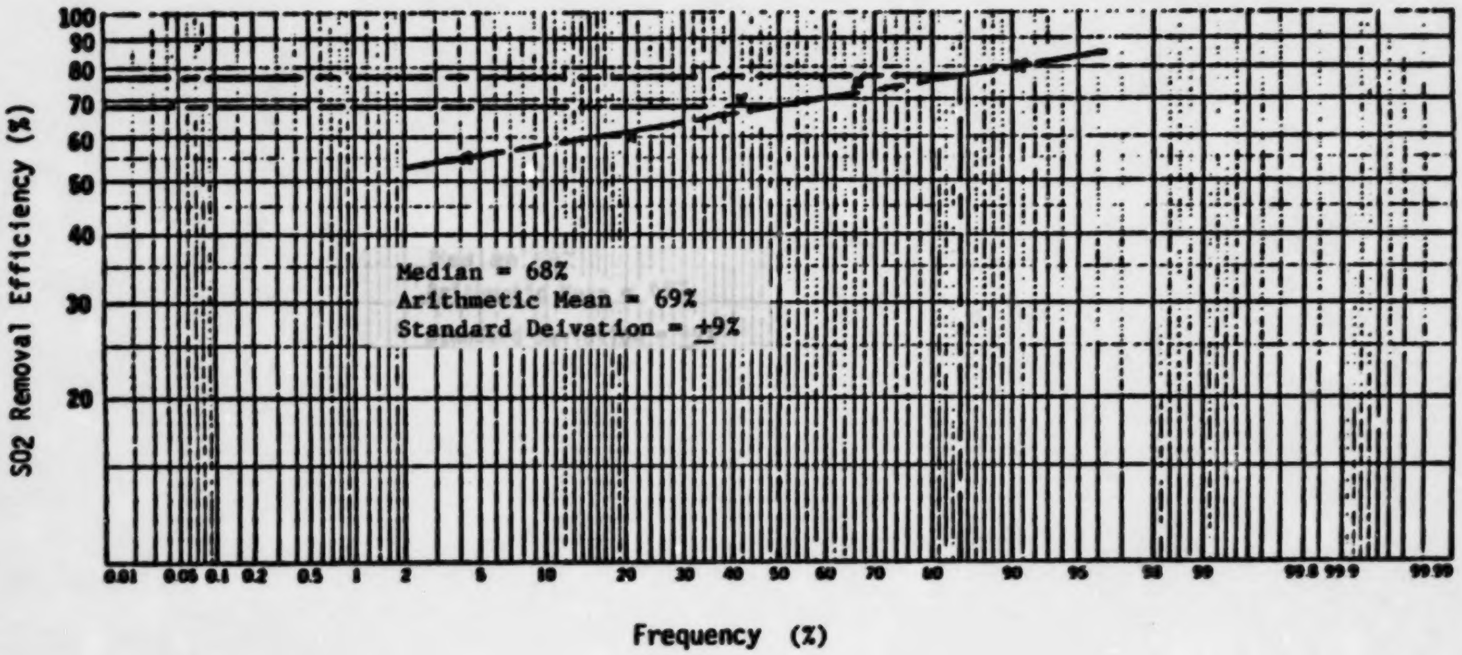

Figure 9. Distribution of SO_2 removal efficiencies.

CONCLUSIONS

Results obtained for SO_2 removal by the spray-dryer/baghouse system treating flue gas from the firing of eastern coal in the spreader-stoker boiler at the Celanese Fibers Company's Amcelle Plant in Cumberland, Maryland, averaged 69 percent, based on 25 days of data over a 33-day test period. During the final days of testing, the removal efficiency was 80 percent, and the system was down less than 2 hours daily.

ACKNOWLEDGEMENTS

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EVALUATION OF EMISSIONS AND CONTROL TECHNOLOGY FOR INDUSTRIAL STOKER BOILERS

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ABSTRACT

This paper presents the results of a 3-phase program to evaluate emissions and control technology for industrial stoker boilers. The paper focuses on the third phase "Limestone/Coal Pellet Development", while summaries are given of the first two phases, "Alternate Fuels Evaluation" and "Control Technology Evaluation". Because SO_2 appears to be the most troublesome emission to control for stokers, a limestone/high sulfur coal pellet was developed and evaluated as a SO_2 control technique. Initially, this pellet with a Ca/S molar ratio of 7 was successfully fired in a 8 MW_{th} industrial spreader-stoker boiler with SO_2 emissions reduced by 75 percent. However, from both an economical and operational standpoint, the amount of limestone required had to be reduced to correspond to Ca/S molar ratio of 3 to 4. Furthermore, the mechanical properties of this pellet were inadequate to withstand the severe stresses of an industrial fuel-handling system. Accordingly, an R&D effort was undertaken to refine the pellet. A refined pellet, with a Ca/S molar ratio of 3-1/2 with appropriate binders was produced that had similar or improved physical characteristics of raw coals. Additionally, economic analysis indicates that this pellet can be produced for approximately $15/ton above the cost of the high sulfur coal. This refined pellet was fired in a 200 kW_{th} laboratory spreader-stoker boiler achieving sulfur captures as high as 70 percent. However, when fired in the 8 MW_{th} (25,000 lb steam/hr) stoker boiler, sulfur captures on the order of 50 percent were achieved.

EVALUATION OF EMISSIONS AND CONTROL TECHNOLOGY FOR INDUSTRIAL STOKER BOILERS

The coal-fired stoker boiler provides an option for industry to meet its energy needs. This option has not been exercised by a significant number of industries primarily because oil- and gas-fired equipment have been, and still are, more environmentally and economically attractive. However, with the dwindling supplies of oil and gas, the rising costs of these fuels, and increased attention given to coal utilization, industry once again is considering the coal-fired stoker boiler.

In support of our nation's commitments to maintain a clean environment and to utilize coal, EPA funded a research and development program to identify and demonstrate improvements in stoker-coal firing that can provide an incentive for greater industrial use of coal. The overall objectives of this program were to

- Characterize the spectrum of emissions from industrial coal-fired stoker boilers using several types of coal under various stoker-firing conditions
- Investigate control methods to reduce these emissions
- Determine the effect of these control methods and variations in stoker-boiler operation on the overall performance of the stoker boiler, and,
- Assess the environmental impact of new technology on the future acceptability of stoker boilers.

This program was recently completed and the final report should be available soon.

This program was divided into three phases. In Phase 1, Alternative Fuels Evaluation, emission characteristics were determined for a variety of coals fired in a 200-kW_t stoker boiler. Emphasis was focused on identifying coals with low pollutant potential, including both physically and chemically treated coals. The results of this phase were presented at the Second Stationary Source Combustion Symposium and contained in its Proceedings.(1) In Phase II, Control Technology Evaluation, potential concepts for control of emissions for full-scale industrial stokers were evaluated. Similarly the results of this phase were presented at the Third Stationary Source Combustion Symposium and contained in its Proceedings.(2) In Phase III, Limestone/Coal Pellet Development, a limestone/coal fuel pellet was developed and evaluated as to its viability as an SO_2 control for industrial stoker boilers. This paper focuses on that effort and summarizes those results.

PELLET DEVELOPMENT

The fuel pellets used in earlier studies(2) did not have adequate strength or durability as up to 50 percent fines were introduced into the $8MW_{th}$ (25,000 lb steam/hr) boiler. Furthermore, from both an economical and operational standpoint, the amount of limestone required to capture a target goal of 70 percent fuel sulfur had to be reduced to correspond to a Ca/S molar ratio of 3 to 4. As a result, an extensive effort (166 test samples) was made to investigate:

- pellet production techniques
- binder types
- coal and limestone particle sites
- limestone types
- pellet formulations.

Laboratory test procedures were developed to evaluate the effect of these variables on the mechanical strength properties of the fuel pellets.

Illinois No. 6 coal was used as the base coal. This coal was ground to 100 percent through 20-mesh and 50 percent through 100-mesh. Ground limestone (-50 mesh) was added to the coal with the selected binders and thoroughly mixed. This mixture was fed to a pellet mill to produce cylindrical pellets one-half inch in diameter and about three-fourth inch long.

Table 1 indicates that pellets were produced with mechanical strength, durability and weatherability characteristics similar to those of raw coals based upon the laboratory test procedures. A number of formulations were identified that could produce satisfactory pellets. The specific formulation used will depend on economics and availability.

LABORATORY EVALUATION

The promising pellet formulations identified during the mechanical strength experiments were evaluated in the model spreader-stoker boiler. This evaluation was based on gaseous emissions (primarily SO_2) and visual observations of the fuel bed. In addition, 18 Mg of the most promising pellet formulation were fired in the Battelle $8MW_t$ steamplant stoker boiler. Criteria pollutants, visual observations, and ash analysis were used in these evaluations.

Model Spreader Experiments

The model spreader-stoker boiler was used to evaluate the more promising pellet formulations. The model spreader provides a simulation of the operation of an industrial stoker boiler and was found to be useful in evaluating the pellets under practical combustion conditions.

Table 2 presents the results of these experiments. In these experiments, the effect of Ca/S ratio (3.5 and 7), the four pellet production techniques, and binder type (cement and methylcellulose) were investigated. Additionally, for comparison, experiments were conducted with medium-S Kentucky coal, Illinois No. 6 coal, and the 50/50 pellets produced during Phase II.

Ca/S Ratio. Bench-scale experiments in a fixed-bed reactor indicated that the Ca/S ratio had little or no effect on sulfur capture for Ca/S ratios greater than 3.5. The model spreader data presented in Table 2 confirm

TABLE 1. COMPARISON OF PHYSICAL PROPERTIES OF RAW COAL AND FUEL PELLETS

Production Method	Pellet Formulation[(a)]				Binder	Durability Index[(b)]	Compression Strength, lb	Weather Index[(b)]	Post Weathering	
	Coal		Limestone							
	Type	%	Type	%					Durability[(b)] Index	Strength, lb
Raw coal	Illinois #6	100	--	--	--	85 ± 2	74 ± 12	89 ± 1	75	58
Raw coal	E. Kentucky	100	--	--	--	85 ± 2	83 ± 22	94 ± 1	83	94
Raw coal	Lignite	100	--	--	--	77 ± 4	92 ± 22	80 ± 4	34	45
Raw coal	Rosebud	100	--	--	--	84 ± 2	50 ± 15	79 ± 2	20	68
CPM lab mill	Illinois #6	70	Piqua	30	2% Allbond + 1% Polyco 2136	87	112	100	85	>112
Banner extrusion	Illinois #6	70	Piqua	30	1.5% Allbond 200 + 1% N-167.01	94	84	100	62	60

(a) Water added as needed.

(b) Percent survival = 100 - percent fines.

TABLE 2. MODEL-SPREADER STUDIES

Run No.	Fuel	Fuel Size, mm	Ca/S Ratio (Approx)	Average SO_2,[a] ppm	Predicted SO_2, ppm	Average Stack Temp, C	Average Excess Air, percent	Average Sulfur Retention, percent	O_2, percent	CO_2, percent	CO, ppm
7-10-78[b]	50/50 CPM pellets (cement)	12.5 x 19	7	854	3700	340	77	77	9.2	10.9	20
8-10-78[b]	100/50 CPM pellets (cement)	Ditto	4	1116	3700	295	85	70	9.7	11.4	—
79-2	50/50 CPM pellets (cement)	Ditto	7	1040	3700	350	140	72	11.5 - 15.0	5.9 - 8.0	—
79-3	70/30 CPM pellets (cement)	Ditto	3.5	1220	3700	360	110	67	10.0 - 13.8	NA	—
79-4	Medium-S Kentucky	—	0	1050	900	—	120	-15	9.5 - 19	8.2 - 10	150
79-5	Illinois #6 coal	—	0	4120	3700	340	95	-11	8.0 - 12.7	8.4 - 10.8	100
79-6	50/50 CPM pellets (cement)	12.5 x 19	7	1240	3700	300	120	67	13.5	8.5	300
79-7	Illinois #6 coal	—	0	3700	3700	325	100	0	10.6	9.2	100
79-8	70/30 CPM pellets (methocel)	12.5 x 19	3.5	1480	3700	335	90	60	9.5 - 11.5	8.0 - 11.2	150
79-9	70/30 briquets (methocel)	12.5 x 25	3.5	1780	3700	365	80	52	8.2 - 9.8	10.8 - 11.6	85
79-10	70/30 disc pellets (methocel)	12.5 dia	3.5	—	—	—	—	—	—	—	—
79-11	70/30 extrusion (methocel)	—	3.5	1370	3700	375	60	63	7.4 - 9.2	10.8 - 12.6	90
79-12	70/30 CPM pellets (methocel)	12.5 x 19	3.5	1260 (1220)[c]	3700	345	75	67	8.4 - 10.2	10.4 - 12	50

(a) Normalized to 3 percent O_2.

(b) 1978 data.

(c) By Method 6.

this observation. Visual observations indicated, as expected, that the pellets with less limestone (Ca/S = 3.5) burned more uniformly and rapidly than those with more limestone (Ca/S = 7).

Production Technique. Pellets using the same formulation consisting of Illinois No. 6 coal, limestone (Ca/S = 3.5), and methylcellulose binder, were prepared by the following production techniques:

- Pellet mill (prepared by Battelle staff)
- Auger extrusion (prepared by Banner Industries)
- Disc agglomeration (prepared by Mars Mineral Corporation)
- Briquets (prepared by Evergreen Company).

The pellet-mill and auger-extruded pellets burned satisfactorily, having sulfur captures of 67 and 63 percent, respectively. The auger extruded pellets were observed to burn more uniformly than the pellet mill pellets perhaps because they were more porous (~1.0 g/cc compared to ~1.4 g/cc).

The briquetted formulations showed relatively low sulfur retention (52 percent) -- a surprising and unexplained result. These pellets burned satisfactorily. The disc agglomerated pellets were entirely unsatisfactory when fired in the model spreader. These pellets disintegrated in the combustion zone producing excessive amounts (greater than 50 percent) of fines. Such fines matted the bed causing nonuniform air distribution. Fuel-bed conditions degraded so rapidly that meaningful data could not be obtained.

Binder Type. Comparison of sulfur retention data of the auger-extruded and mill-pellets made with organic (methylcellulose) and inorganic (cement) binders indicated no significant difference. The binders are used in very small quantities (less than 4 percent) and do not have any catalytic effects. As a result, it appears that the type of binder does not significantly affect the combustion behavior of the pellet providing the physical properties of the pellet are retained. Cement-bound pellets with satisfactory physical properties could not be made by the disc agglomeration and briquetting methods.

Steamplant Stoker Demonstration

Eighteen Mg of the limestone coal fuel pellets with a Ca/S molal ratio of approximately 3.5 were fired in the $8MW_t$ steamplant boiler. Two types of pellets were used — a lower density (0.9 to 1.2 g/cc) pellet produced by Banner Industries using sugar extrusion and a higher density pellet ($\simeq$ 1.4 g/cc) produced by Alley-Cassetty Coal Company using a pellet mill. Both types of pellets were fired under a variety of boiler conditions. Evaluations were based on visual observations, criteria pollutants, and ash analyses.

Pellet Formulation

Allbond-200 cornstarch and M-167 latex emulsion were used as binders. The resulting pellet formulation (dry basis) consisted of:

67 percent Illinois No. 6 coal
30 percent Piqua limestone
2 percent Allbond 200 binder
1 percent M-167 latex binder.

The pellets remained sufficiently intact during storage and handling that an acceptable pellet was fed into the boiler. However, it was observed that some pellets softened during exposure to rain. Weatherability tests on these pellets were rerun showing approximately the same characteristics. It appears that the weatherability test used during pellet development has some limitations and that pellets will require some undercover storage or further formulation refinement for weatherproofing.

EXPERIMENTAL RESULTS

Checkout Runs. Prior to the demonstration test, the fuel pellets were fired for 10 hours to determine the necessary stoker adjustments and to establish a range of operating conditions.

The stoker feed mechanism distributed the pellets uniformly over the grate. This was unexpected since the pellets were all approximately the same size. It was observed, however, that approximately 50 percent of the pellets broke randomly into smaller pieces providing a reasonably good size distrubution.

a. Phase II/Phase III Pellet Comparison. Pellets fired in the Phase III study were significantly superior to those fired previously in the Phase II steamplant runs. They burned more readily at lower excess air rates, provided improved boiler response (thinner bed), ignited more readily, and generated lower CO and smoke levels. These improvements are attributed to the fact that the Phase III fuel had a higher heating value, contained an organic (rather than inorganic) binder, contained less ash, and exhibited superior mechanical strength. However, sulfur retention was not as high with the Phase III pellets.

b. Stoker Coal/Phase III Pellet Comparison. Phase III pellets appeared to burn equally as well as the low-sulfur Ohio coal that is normally fired in the Battelle steamplant boiler. The boiler appeared to be as responsive to the load and could be operated at comparable excess air levels. Table 3 compares these two fuels. Emissions are corrected to 3 percent O_2.

c. Effect of Operating Parameters on Sulfur Retention. Because it was not the intent of the checkout runs to characterize the emissions for a variety of boiler operating conditions nor was it possible with the limited supply of fuel pellets, only limited amounts of data were collected in the checkout runs.

Sulfur retention was observed to decrease for increasing load as indicated below for relatively constant excess air (about 80 percent).

TABLE 3. COMPARISON OF EMISSIONS FROM COMBUSTION OF A LOW SULFUR COAL AND LIMESTONE/COAL PELLET

Coal Type	Smoke Opacity, percent	CO	NO	Fuel N Converted, percent	SO_2	Fuel S Emitted, percent
Low-S coal	10	70	480	18	540	90
Fuel pellet	20	400	310	20	1800	45

Boiler Load, percent full load	Sulfur Retention, percent	Bed Temperature, C
0.64	50	1315
0.80	48	1405
0.85	47	1425

The bed temperatures were measured with an optical pyrometer sighted on the combustion zone at the top surface of the bed. Sulfur retention varies with bed temperature. However, this observation must be tempered as the combustion conditions were not closely controlled throughout these and the observed temperature measurement may not be a good indication of the actual bed temperature.

At a low-load condition, the excess O_2 was varied from 9.5 percent to 16 percent with no significant change in the SO_2 retention (46 to 50 percent). Bed depths were also varied from 80 to 160 mm. SO_2 retention increased somewhat with deeper beds. The increased retention was attributed to the lower bed temperatures measured for the deeper beds.

DEMONSTRATION TEST

During the limestone/coal fuel pellet demonstration, the pellet feed rate was maintained at approximately 1360 Kg/hr for a boiler load of 80 percent. Tables 4, 5, 6, and 7 summarize the results of this test.

a. Sulfur Capture. As indicated in Table 4, sulfur capture was 45 percent during the demonstration test. This sulfur retention is less than that observed for the model spreader and fixed-bed reactor experiments firing pellets of similar formulations. Additionally, as previously discussed, a 75 percent sulfur retention was achieved when firing a cement-bound pellet with a Ca/S ratio of 7 in the steamplant during Phase II. The greater sulfur retention of these other experiments is attributed to the lower bed temperatures, which seldom exceeded 1260 C. The bed temperatures in the Phase III steamplant demonstration were seldom less than 1370 C and ran as high as 1455 C. Additionally, with a pulsating

TABLE 4. EMISSION DATA SUMMARY FOR FUEL PELLET DEMONSTRATION

Load, %	O_2, %	CO_2, %	CO, %	NO, %	SO_2, %	Smoke Opacity, %	CO at 3% O_2, ppm	NO at 3% O_2, ppm		Fuel N Converted, %	SO_2 at 3% O_2, ppm		Fuel S Emitted, %	Particulates, ng/J
								Computed	Measured		Computed	Measured		
80	8.4	10.5	300	310	1600	20	420	2250	440	20	4100	2250	55	258

TABLE 5. ANALYSIS OF METHOD 5 FILTER CATCH (Weight Percent)

Ash	C	Ca	CO_3	Fe	Total S
81	19	11	--	4	54

TABLE 6. ANALYSIS OF GRATE DISCHARGE (Weight Percent)

Ash	C	Ca	CO_3	Fe	Total S
97.7	1.8	36.5	0.7	5.8	3.9

TABLE 7. SULFUR BALANCE

Computed Fuel S In, lb/10^6 Btu	Emitted as SO_2, lb/10^6 Btu	Sulfur Retained in Bed Ash as SO_2, lb/10^6 Btu
7.4 (3182 ng/J)	4.1 (1763 ng/J)	3.3 (1419 ng/J)

ash discharge stoker, the fuel bed is violently disturbed. Ash can be recirculated back into the hot zone. Thus, if sulfur is retained in the ash at a lower bed temperature, it may be released when the ash is exposed to a higher temperature.

The average SO_2 emission level of 1600 ppm during the Method 5 test was verified by the Method 6 wet-chemistry technique. (Wet chemistry gave an SO_2 emission level of 1590 ppm.) In addition, as indicated in Table 7, the sulfur balance based on the fuel pellet analysis, the SO_2 emission and the sulfur content in the bottom ash (Table 6) was complete.

b. CO Levels. CO levels from pellet firing were relatively high compared to those from the firing of conventional stoker coals which are usually <100 ppm. These higher CO levels may be related to the

A possible explanation for the higher CO levels was that the overfire air rate was significantly decreased during pellet firing. In the Battelle boiler the overfire air jets are only 250 mm above the grate. With the increased bed depth from pellet firing, the overfire air jets would have impinged upon the fuel bed if the normal flow rate were maintained. The impingement would increase ash carryover, increasing particulate loadings.

c. Particulate Loading. The Battelle steamplant boiler facility has a mechanical collector to control particulates. Depending on the ash and sulfur content of the coal, the experiments in Phase II showed that particulate loadings varied between 86 and 258 ng/J (0.2 and 0.6 $lb/10^6$ Btu). Generally, for low S, low ash coals, particulate loadings were less than 129 ng/J (0.3 $lb/10^6$ Btu).

The particulate loading from the firing of the fuel pellet was 258 ng/J (0.6 $lb/10^6$ Btu). This loading was not unusually high for a spreader stoker firing a 33-percent-ash coal. This loading should be significantly less for a chain-grate stoker. The smoke opacity was only 20 percent, which would appear low for a particulate loading of 258 ng/J if the fly ash collected was from conventional stoker coal. However, the fly ash from pellet firing is about 50 percent more dense and considerably more coarse than from conventional coals. For equivalent mass loadings, optical density varies inversely with particle size and density. Thus, the apparent discrepancy between smoke opacity and particulate loading is explained partially by laws of optics. As indicated in Table 5, about 19 percent of the fly ash was carbon, a negligible carbon loss.

d. Grate Discharge. Table 7 shows that the unburned carbon content in the grate discharge was less than 2 percent. This indicates that the fuel pellets were burned essentially to completion. Analysis indicates that Ca and SO_4 were present and could have combined with water to form a solid mass. Some minor plugging problems were experienced in the ash-disposal system when steam was used to control dusting during transport of the ash.

SUMMARY

The steamplant demonstration indicated the limestone/coal fuel pellet could be fired in an acceptable manner without modifying the facility. During the demonstration, sulfur capture levels that would make the fuel pellet a viable SO_2 control were not achieved. The data suggest that improved SO_2 retention could be realized if bed temperature could be reduced to below 1315 C, perhaps with flue gas recirculation. In addition, a quiescent fuel bed in a stoker boiler may increase the sulfur retention in the bed and should reduce particulate emissions.

LIMESTONE/COAL FUEL PELLET PROCESS COST SUMMARY

Table 8 summarizes an economic analysis of the limestone/coal pellet process. This analysis considers costs related to raw materials, utilities, labor, and capital, including profit, interest, and income tax. It indicates a process cost of approximately $15.40/Mg ($14/ton) of pellets in addition to the cost of the high sulfur coal. Increased costs of firing the boiler are not considered. As an example of such costs, because of the high ash content of the pellet, ash handling and disposal costs would be higher than for the low-ash conventional coals.

TABLE 8. SUMMARY OF LIMESTONE/COAL PELLETIZING PROCESS COSTS

Basis: 60 tons per hour product with 65 percent coal, 30 percent limestone, 5 percent Portland cement
23 hours per day, 330 days per year
1380 tons per day, 455,400 per year of product

Fixed plant investment	$2,790,000
Working capital	80,000
Interest during construction	250,000
	$3,120,000

Item	Annual Costs, Dollar	Per Ton Product, Dollars
Raw Materials		
Limestone 18 tons/hr, 136,620 tons/year at $8/ton delivered	$1,092,960	2.4
Pregelatin cornstarch, 9100 tons/year at $20/ton delivered	1,138,500	3.0
Latex emulsion, 1.2 ton/hr, 9100 ton year at $150/ton delivered		
Utilities		
Process water 12 tons/hr (48 gpm) 21.9 MM gallon/year at $0.2/M gal	4,400	0.01
Fuel oil 32 MMBtu/hr, 243 trillion Btu/yr of $3/MMBtu	728,600	1.60
Power 75 percent of 1917 KW or 1440 KW at $0.035/KW-hr	382,500	0.84
Diesel fuel 5 gph, 37,950 gallon/year at $0.80/gal	22,800	0.06
Labor Related		
Direct labor — 7 operators @ $8/hr plus 25 percent payroll burden ($10/hr total); staffed 365 days/yr	613,200	1.30
Supervision — 15 percent of direct labor	91,980	.20
Overhead — 50 percent of direct labor and supervision	306,500	.70
Capital Related		
Maintenance — 6 percent of fixed plant investment	167,400	0.37
Special pelletizer maintenance at $0.30/ton plus $0.55/ton die and rollers	387,100	0.85
Front-end loader maintenance at $0.22/hr per machine	3,300	0.01
Taxes and insurance — 1.5 percent of fixed plant investment	41,850	0.09
Depreciation — 11 year, straight line on fixed plant investment	250,000	0.56
Profit, interest, income tax — 30 percent of total employed capital	936,000	2.05
TOTAL	$6,167,100	~$14.00

The estimated cost of \$15.40/Mg of pellets above the price of the raw coal is based on the best available data. The cost may vary depending on the type of system used and whether the process may be integrated into a physical coal cleaning preparation plant. This cost is for a product with a heating value of 18.6 KJ/g (8000 Btu/lb) and thus adds about \$0.95 per 10^9 joules (\$1 per million Btu) for SO_2 control. It indicates that the limestone/coal pellet is cost competitive with other control strategies.

BASIC ASSUMPTIONS

The following assumptions were used in the analysis.

- Mine-mouth operation
- Limestone and coal ground to 60 to 100 mesh
- Pellet composition:

  65 percent high sulfur coal
  32 percent limestone
  2 percent pregelenized cornstarch
  1 percent latex emulsion

- Plant capacity of 54.4 Mg/hr (60 tons/hr).

The pellet composition was based on the results of the pellet development effort.

PROCESS FLOWSHEET

The economic analysis was based on the process flowsheet presented in Figure 1. In this process

- Coal is taken from a pile instead of directly from an existing mine operation conveyor
- Limestone is delivered to a pile by truck
- Portland cement is delivered directly to a bin from a truck by pneumatic feeding system suggested by Jeffrey Manufacturing Company
- Relatively long inclined conveyors from the coal and limestone piles are assumed. Costs would be about 35 percent less for horizontal conveyors combined with bucket elevators.

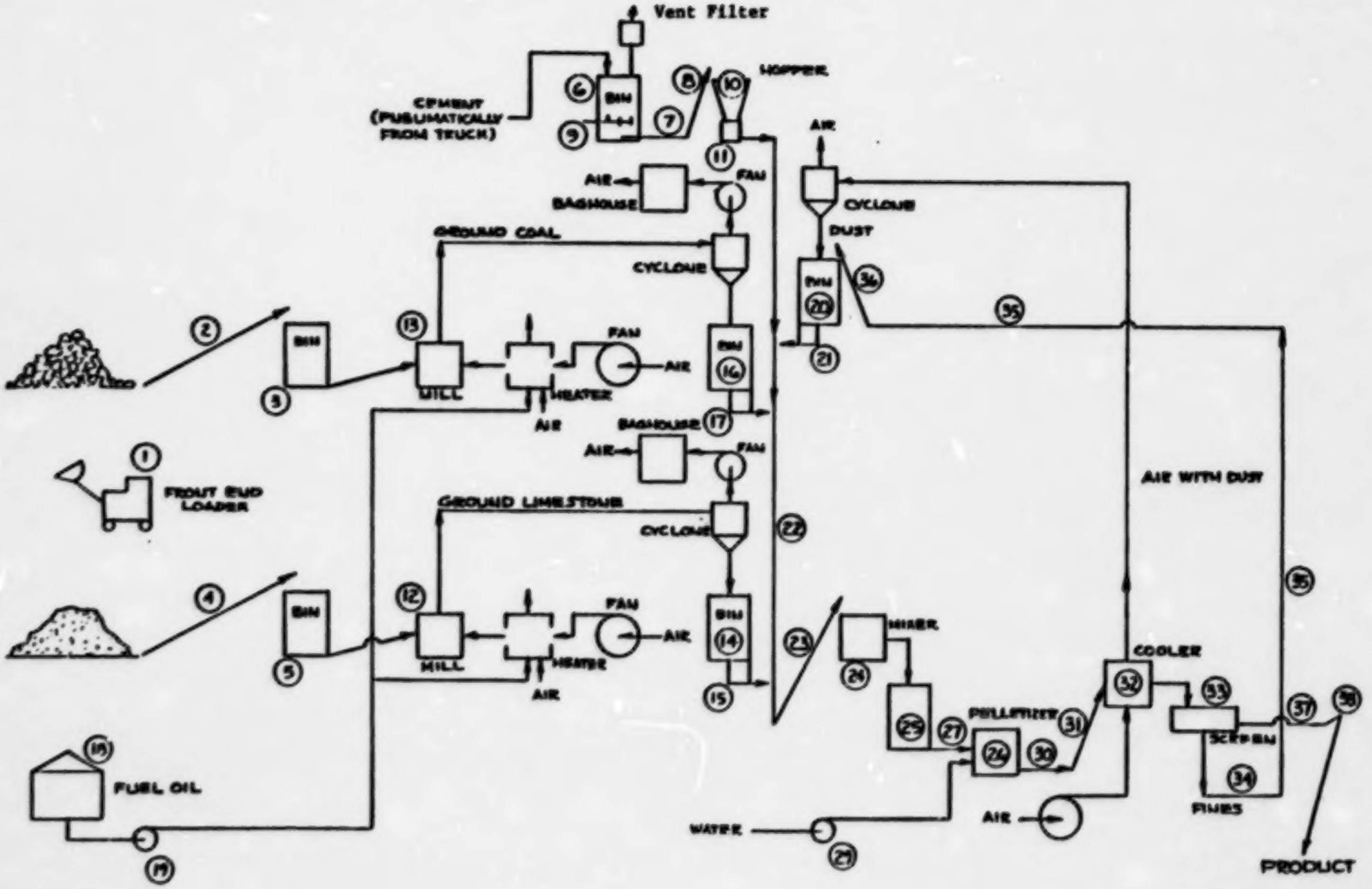
Vent Filter
CEMENT (PNEUMATICALLY FROM TRUCK)
6
BIN
9
7
8
10
HOPPER
11
AIR
BAGHOUSE
FAN
CYCLONE
DUST
20
BIN
21
34
35
GROUND COAL
2
BIN
3
13
MILL
HEATER
AIR
FAN
AIR
BIN
16
17
1
FRONT END LOADER
BAGHOUSE
AIR
FAN
GROUND LIMESTONE
CYCLONE
22
AIR WITH DUST
4
BIN
5
12
MILL
HEATER
AIR
FAN
AIR
BIN
14
15
23
MIXER
24
25
27
PELLETIZER
26
30
31
COOLER
32
33
SCREEN
34
FINES
35
37
38
PRODUCT
18
FUEL OIL
19
WATER
28
AIR

FIGURE 1. COAL/LIMESTONE/CEMENT PELLETIZING PROCESS FLOWSHEET

- A paddle-type mixer, as suggested by California Pellet Mill, is used
- California Pellet Mill pelletizers and dryers are costed.

A California Pellet Mill was used in the analysis since cost information was available. However, pellets can be produced by an extruder at perhaps a lower cost. Specifications for processing equipment are given in Table 8.

SOURCES OF INFORMATION

Information on equipment included in the flowsheet was obtained from the following sources:

Front-end loaders — Caterpillar Tractor
Conveyors/elevators -- Jeffrey Manufacturing
Storage bins — Butler Manufacturing
Feeders — Jeffrey Manufacturing
Solids mixer — Rapids Machinery
Pelletizers -- California Pellet Mill
Coolers — California Pellet Mill

COMPARISON TO OTHER CONTROL STRATEGIES

The limestone/coal fuel pellet is an attractive control for two major reasons:

(1) No major modification of the stoker boiler facility is required to fire the pellets

(2) The cost of \$15.40/Mg is competitive with other control strategies such as used flue gas scrubbers or low sulfur coals.

The steamplant experiments indicate that neither the stoker boiler facility nor its operation will require major modification to fire fuel pellets. The pellets burn similarly to a lower heating value coal. In contrast, the addition of a flue gas scrubber is a major facility modification and increases system maintenance.

Cost comparisons of the various types of control strategies are difficult to interpret, primarily because of different sets of basic assumptions and different reference points. However, the pellet process costs of \$15.40/Mg or \$0.95 per 10^9 joules (\$1 per million Btu) are competitive with flue gas scrubbers. Foley (3) indicated costs of between \$22 and \$33/Mg (\$20 and \$30/ton) of coal for the gas scrubber for small to medium-sized industrial boilers based on 1973 figures.

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UNPRESENTED PAPERS

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FLAKT'S DRY FGD TECHNOLOGY: CAPABILITIES AND EXPERIENCE

BY

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ABSTRACT

Dry FGD technology has rapidly developed in the past few years to the extent that it is now the predominant FGD process for low-sulfur coal applications.

This paper describes two pilot demonstration projects using Flakt's dual-fluid nozzle atomizer and both an electrostatic precipitator and fabric filter. The design, operation, and test results from pilot installations in Denmark and the United States are reviewed. The significant contribution to SO_2 removal of the fabric filter, flyash "precollection," and product recirculation are discussed.

INTRODUCTION

Dry flue gas desulfurization (FGD) has rapidly expanded in the past three years as an economical means of SO_2 abatement for low-sulfur coal applications. Its rapid development from a few small-scale pilot plants to full-scale contracts for a total of 3500 MW without a single prototype demonstration plant is remarkable. Unlike the first commercial applications of wet lime/limestone technology in the early 70's dry FGD is often said to be merely a new combination of several proven technologies.

Although the current process concept was suggested as early as 25 years ago in a Czecholovakian patent, the recent impetus to exploit Western low-sulfur coal reserves has stimulated the current explosion of dry FGD technology. Dry FGD is traditionally considered as two separate techologies: spray drying and particulate collection. An alkali absorbent reacts with SO_2 in a spray dryer where a dry reaction product is formed. This reaction product together with flyash is then collected in a conventional dust collector which is usually a fabric filter. From a commercial standpoint these two technologies have been developed and supplied by two separate companies joined together under various commercial arrangements.

It is Flakt's contention that dry FGD must be approached like every other FGD technology -- as a complete integrated process system. Optimization of this process requires tradeoffs between the dry scrubber ("spray dryer") and dust collector as well as the ancillary equipment and control system.

The dry scrubber must be considered as a chemical reactor wherein the SO_2 reacts with an alkali absorbent to form certain reaction products. The operating process parameters must be optimized with the objective to maximize the SO_2 mass transfer rate. This maximization, however, is constrained by the essential requirement to maintain a "dry bottom," i.e., dry reaction prducts, at all times. This requirement is a limitation to the process variables which control SO_2 mass transfer. It is not the primary objective as it is with conventional spray drying technology where the dry product quality is most important. For this reason, many of the operating parameters and design characteristics that are appropriate to spray drying are not required nor even desired for SO_2 removal.

The predominance of the fabric filter as the desired particulate collector is a result not only of its high dust collection capability, but also of its demonstrated ability to absorb residual SO_2. The use of the fabric filter as a chemical reactor is well-known and it is this secondary absorption ability that must be integrated with the dry scrubber's performance to optimize a dry FGD system. As discussed herein, Flakt's pilot testing has indicated that 30-40% of the total SO_2 removal can occur within the fabric filter.

The dry FGD process is a system with several interactive elements and one cannot ignore the importance of the ancillary equipment required for alkali preparation and waste disposal. The nature of the dry FGD process dictates the design of these ancillary systems which also must be integrated together.

Lastly, one of the most critical elements that must be considered is the process control system. It is essential that the control system respond to the requirements of the FGD process with recognition of the operating limitations of the equipment.

Flakt's dry FGD technology is the product of drawing upon over 25 years of experience worldwide in the background technologies and integrating these throughout a methodical development program. Flakt's current dry FGD development started in 1978 with bench-scale investigations of the chemical thermodynamics and kinetics associated with these rather complex gas-liquid-solid reactions. This bench-scale work continues today.

Flakt's development has reached the stage where two field plants are now operating in Denmark and the United States. A commercial-scale demonstration plant is also contemplated for which plans will be announced shortly.

This paper will review Flakt's experience in the design, operation, and test results from the two field pilot plants.

ASNAES FIELD PILOT PLANT

Plant Description

The Asnaes Power Station is located at Kalundborg on the west coast of Zealand in Denmark. It is owned by IFV (Elektricitetsselskabet Isefjordsvaerket I/S) and in 1980 it will become the largest coal-fired power station in Northern Europe with five boilers with a total generating capacity of 1250 MW.

Coal is imported by ship from different sources and the quality of the coal will vary within wide limits.

This situation is expected to exist for most future power stations in Scandinavia, therefore, the pilot plant test conditions are representative of existing and future installations. Any flue gas desulfurization system at a coal-fired power station in Scandinavia must be designed with sufficient flexibility to allow efficient operation for varying coal sources. These particular aspects were carefully considered when designing the Flakt pilot plant which is located on Unit No. 4 at Asnaesverket, a 250 MW pulverized coal-fired boiler.

Pilot Plant Description (See Figure 1)

A slip stream of 10,000 acfm is isokinetically extracted from a duct upstream of the electrostatic precipitators. The gas flow can then either pass through a cyclone where a major portion of the flyash is collected or fed directly into the SO_2 reactor where the absorbent lime slurry is sprayed into the hot flue gas by means of a dual-fluid nozzle.

Before the gas is returned to the main flue gas duct, it passes through a high-ratio fabric filter where the flyash and reaction products are collected. A significant portion of the total SO_2 absorption occurs in the fabric filter where the SO_2 contacts the dust deposited on the bags.

Lime slurry dilution water to achieve proper reactor outlet temperatures, and - where applicable - recycled material (flyash and absorption products) are added to the absorbent feed tank.

The lime input to the system is proportional to the SO_2 concentration in the flue gas, whereas the dilution water flow is contolled via a simple level switch. The reactor outlet temperature is by far the most important process parameter. It is controlled by a variable speed pump which allows excellent temperature control by varying the slurry flow to the nozzle.

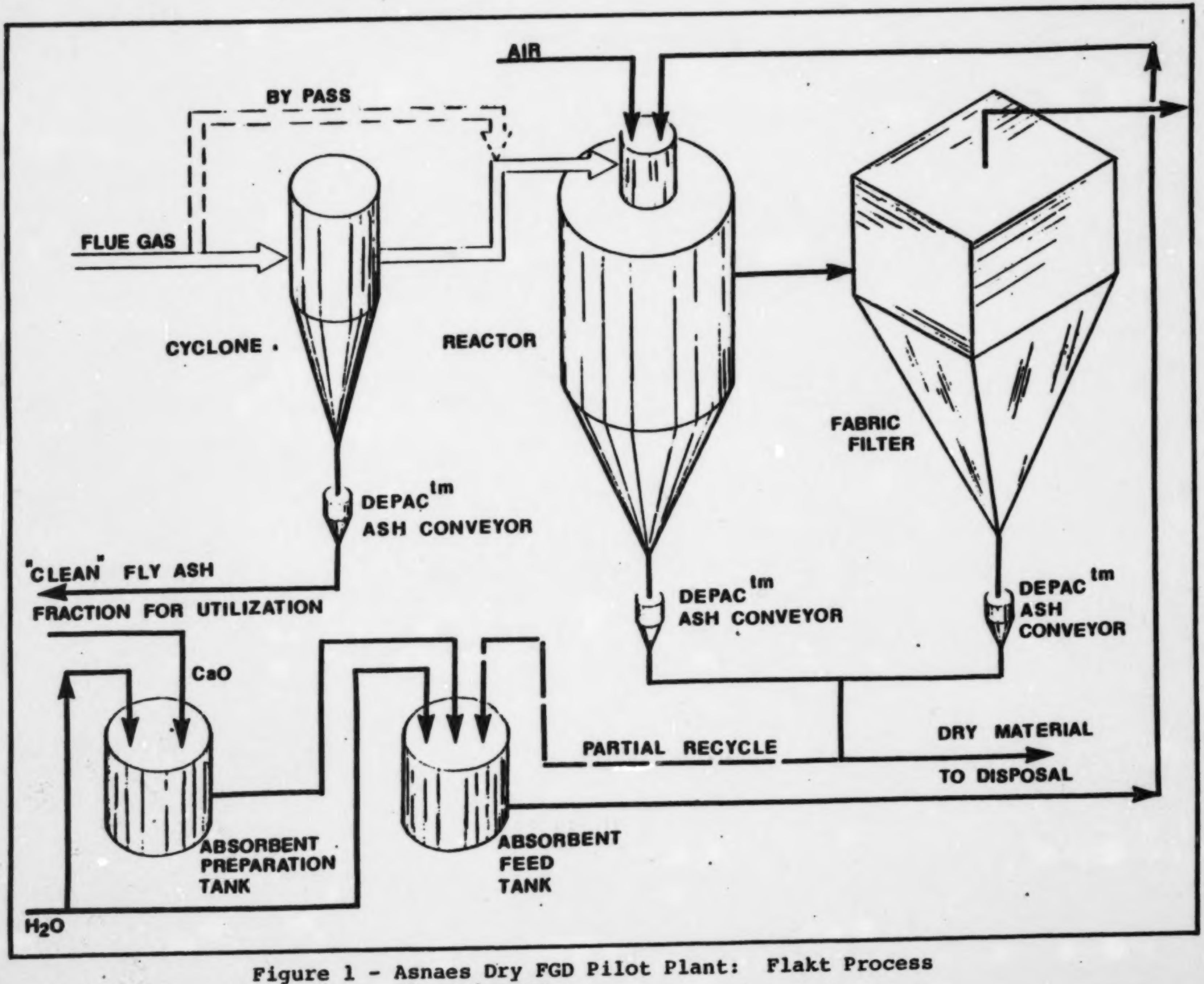
AIR
BY PASS
FLUE GAS
CYCLONE
REACTOR
FABRIC FILTER
DEPAC^{tm} ASH CONVEYOR
"CLEAN" FLY ASH
FRACTION FOR UTILIZATION
DEPAC^{tm} ASH CONVEYOR
DEPAC^{tm} ASH CONVEYOR
CaO
PARTIAL RECYCLE
DRY MATERIAL
TO DISPOSAL
ABSORBENT PREPARATION TANK
ABSORBENT FEED TANK
H$_2$O

Figure 1 - Asnaes Dry FGD Pilot Plant: Flakt Process Flow Diagram

The fabric filter used at Asnaes is a Flakt type OPTIPULSE$^{(R)}$. This design is in successful operation at coal-fired power stations in Europe and North America, e.g., the 150 MW installation at the H. R. Milner Power Station owned by Alberta Power, Canada.

The pilot filter contains 72 bags arranged in 6 rows. Each row is equipped with a separate pulse valve. When actuated, 12 bags are cleaned at a time.

The operation of a fabric filter in connection with a dry scrubber is somewhat different compared to operation only for dust collection. A "clean" filter bag does not contribute to the SO_2 absorption until a dust cake is formed on its surface and thus the total SO_2 absorption efficiency follows the cleaning cycle of the fabric filter. To increase the probability of reaction between SO_2 and the dust cake, the amount of material present on the bags should be maximized. The fabric filter is, therefore, operated at a constant differential pressure and just one pulse valve is actuated at the same time. Each pulse typically increases SO_2 emission by 20 - 30 ppm over a short period. Applied to a large filter, however, the difference in performance when pulsing will be negligible.

Coal and Flyash Properties

The coal types tested at Asnaes are listed below in Table 1. The sulfur content in the coals corresponds to 500 - 1250 ppm SO_2 in the flue gas.

Table 1

Source	Heat Value BTU/lb	Ash %	Sulfur %
Poland/W.Canada (mix)	10275	15.4	0.8
USA	10955	15.5	1.2
S. Africa	10570	14.7	0.6
USA	10630	15.6	0.8
Poland	9880	17.2	0.4
USA	9520	23.3	0.3
Poland	9885	16.7	0.8

The flyash alkalinity contributes to the overall SO_2 removal efficiency; the higher the alkalinity is, the higher the expected contribution. The flyash alkalinity of the coals burned at Asnaes varied between 0.4 and 2.2 moles/kgs or 0.08 - 0.72 moles/mole SO_2 (defined as flyash alkalinity in flue gas entering reactor per mole of SO_2 entering reactor). These low alkalinites should be observed, since the situation may be quite different for other types of coal. For example, alkalinity in flyash from North Dakota lignite is in the range of 9 - 12 moles/kg corresponding to 4 - 6 moles/mole SO_2.

Test Results - Once-Through Operation

Within the range of coals studied, results from once-through operation have been remarkably uniform despite different SO_2 contents, flyash alkalinities, etc. This indicates that in this mode of operation the contribution from the flyash alkalinity to SO_2 absorption is marginal. Only minor changes in SO_2 efficiency were recorded when switching from operation with or without the flyash pre-collector, which further verifies that the SO_2 reactive alkalinity was essentially derived from the absorbent, see Figure 2.

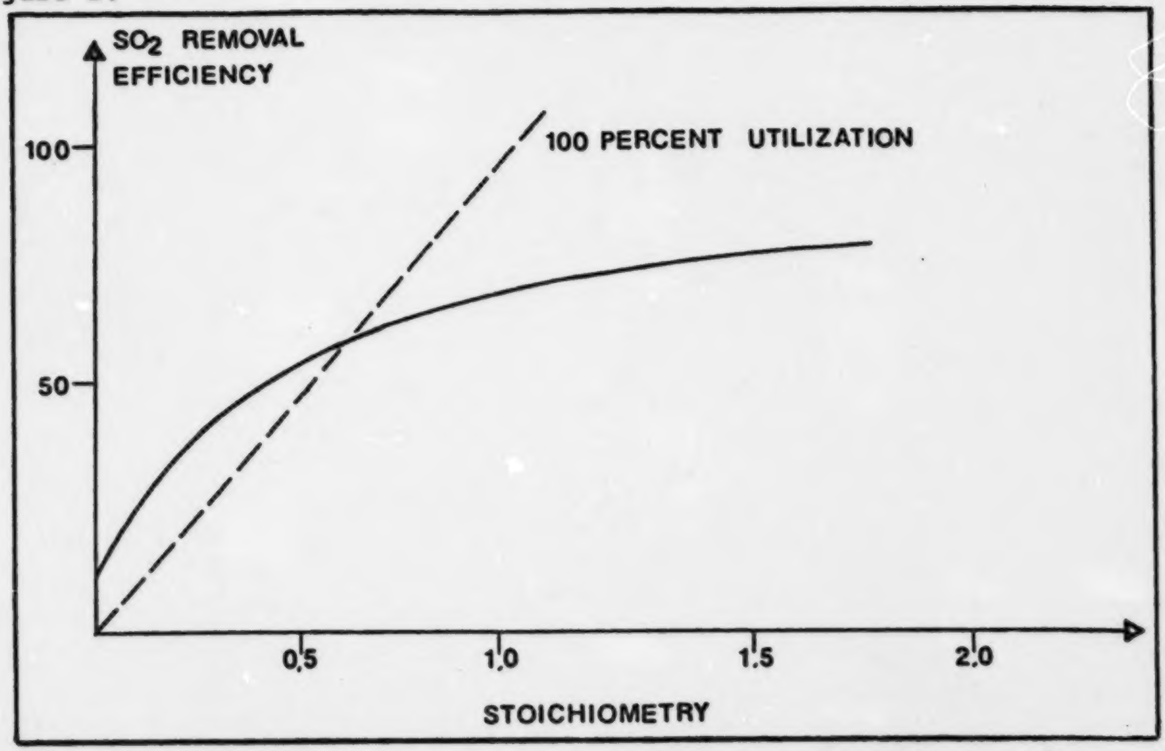

Figure 2: SO_2 removal performance as a function of lime stoichiometry - once-through operation.

However, at low stochiometries (<0.5) utilization of the absorbent generally was about 100%, which demonstrates that the ash in these cases contributed to the SO_2 absorption. When firing a certain type of Polish coals and spraying only water into the reactor, SO_2 removal efficiency was 12%.

The similarity with wet systems utilizing flyash as the only absorbent is striking, as in this case operation at low pH is essential to promote dissolution of alkali from the flyash. A low excess of the absorbent apparently corresponds to operation at a low pH value in a wet process.

Generally speaking, SO_2 absorption is favored by a low outlet temperature - or rather a high humidity - and if no other restrictions were necessary, the ideal process would mean operation very close to or at the saturation temperature. However, there seems to be an equilibrium between water vapor in the flue gas and the mixture of reaction products/flyash. Therefore, the dust gradually becomes somewhat moist at temperatures 10 - 15°F above the saturation temperature. Hence, operation at temperatures closer to the saturation temperature seems virtually impossible for a commercial installation.

Recycle Operation

Recirculation of the absorbent is a standard procedure for wet SO_2 processes and was already suggested in 1958 in connection with dry scrubbing (Czechoslovakian Patent No. 96138). The recirculation scheme for a dry system is somewhat more complicated as it requires handling of relatively large quantities of dust and/or slurry as well as an increased potential for scaling and plugging in the wet parts of the system.

An important factor to consider is the highly abrasive nature of the flyash which is added to the lime slurry. The flyash is more abrasive than the lime and hence causes additional wear on pipes, pump linings, etc. Flakt proprietary dual-fluid nozzles are designed to handle slurries. The nozzles are equipped with inserts, made of a specially selected material with mechanical and abrasion resistant properties superior to cemented carbides, conventional ceramics or even silicon carbide.

The results of the recycle tests are very informative as they point out one possible way to enchance the performance of a dry scrubber even if the flyash is low in alkalinity.

Considering the fact that aluminates and silicates, the major constituents of flyash, are inert with respect to SO_2, it was shown that a pre-collector for flyash will in fact "enrich" the recirculating material with respect to the alkali absorbent and its reaction products. The higher the removal efficiency of the pre-collector is, the more of the absorbent that can be recirculated to the reactor.

The advantage of recirculation with pre-collection of flyash is illustrated in Figure 3. Two runs, with and without the operation of the pre-collector for flyash, are shown. About 80% SO_2 removal efficiency is possible to achieve at nearly complete utilization of the lime added.

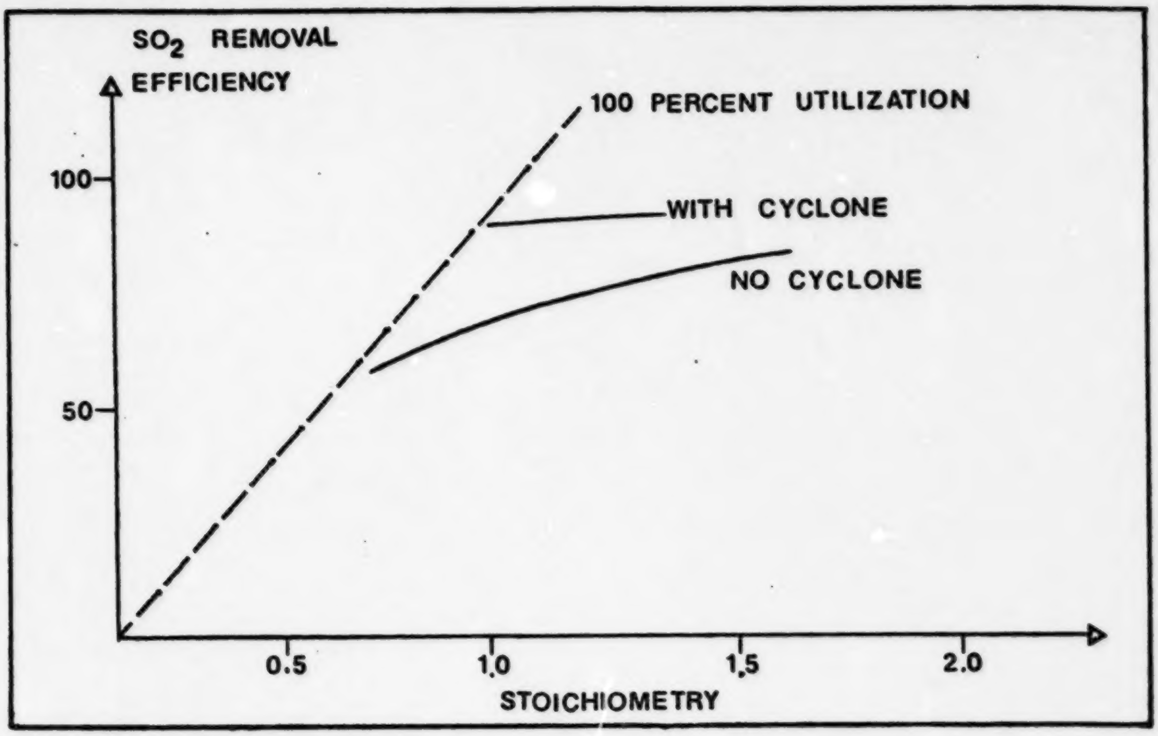

Figure 3: SO_2 removal efficiency as a function of lime stoichiometry, and demonstrating the effect of precollection and recirculation.

The cost for a flyash pre-collector, i.e., one-field EP, can thus be justified, due to the savings in reagent cost.

Additional advantages of a flyash pre-collector are:

. less wear, due to a low flyash content in the recycle slurry; furthermore, most of the troublesome coarse fraction of the flyash is removed in the pre-collector.

- flyash removed in the pre-collector is not contaminated with reaction products.

The fact that a "clean" fraction of flyash can be recovered from the dry scrubbing system is very important in those areas where the ash is sold, e.g., for use in cement or concrete. It is estimated that 6,300,000 tons of flyash was utilized in the U.S. in 1977.

Dry scrubbing in particular requires reliable transportation and handling of flyash/reaction products. To demonstrate the performance of the Flakt dense phase pneumatic conveying system (DEPAC^TM), the cyclone, reactor and the fabric filter have been equipped with dust transmitters.

JIM BRIDGER PROJECT

The Jim Bridger Electric Generating Station, located some 35 miles northeast of Rock Springs, Wyoming, and jointly owned by the Pacific Power and Light Company and Idaho Power Company, has been the site of Flakt's most comprehensive dry FGD pilot program.

Power Plant Description

The Jim Bridger Plant is capable of generating 2000 MW and is one of the largest single sources of electric power in the Rocky Mountain region. The fuel source for the four generating units is the Bridger Mine, which produces coal from a large coal field adjacent to the power plant. This coal is ranked as sub-bituminous in quality with an average thermal value of 9600 Btu/lb, and contains 10 percent ash and an average of 0.6 percent sulfur.

The steam generators for the four units are manufactured by Combustion Engineering Co. The flue gas is treated for particulate removal in Flakt electrostatic precipitators installed on all units at the plant.

Pilot Plant Description (See Figure 4)

The conceptual design for the pilot plant was started in April of 1979. Since Flakt had already done extensive work on a bench-scale, together with development work on gas/liquid droplet mixing techniques, it was important that the Jim Bridger Pilot Plant exhibit features which, in addition to testing these concepts, allowed for the study of other aspects of an integrated FGD system.

The pilot plant includes two alternative reactor configurations coupled with two alternative particulate removal devices (i.e., electrostatic precipitator and fabric filter).

A flue gas slip stream of 15000 ACFM is isokinetically extracted from an existing EP inlet duct. This flue gas, which is normally at a relatively low temperature (220-250°F), then enters either of the two reactors.

Since layout restrictions may be more limiting than process requirements at some locations, both a vertical and a horizontal tower reactor have been installed. The vertical tower is 8½ ft in diameter and approximately 70 ft high. There is a coned bottom with a rotary valve to allow removal of dust fallout from the tower. The horizontal tower is 9 ft x 6½ ft and 30 ft long

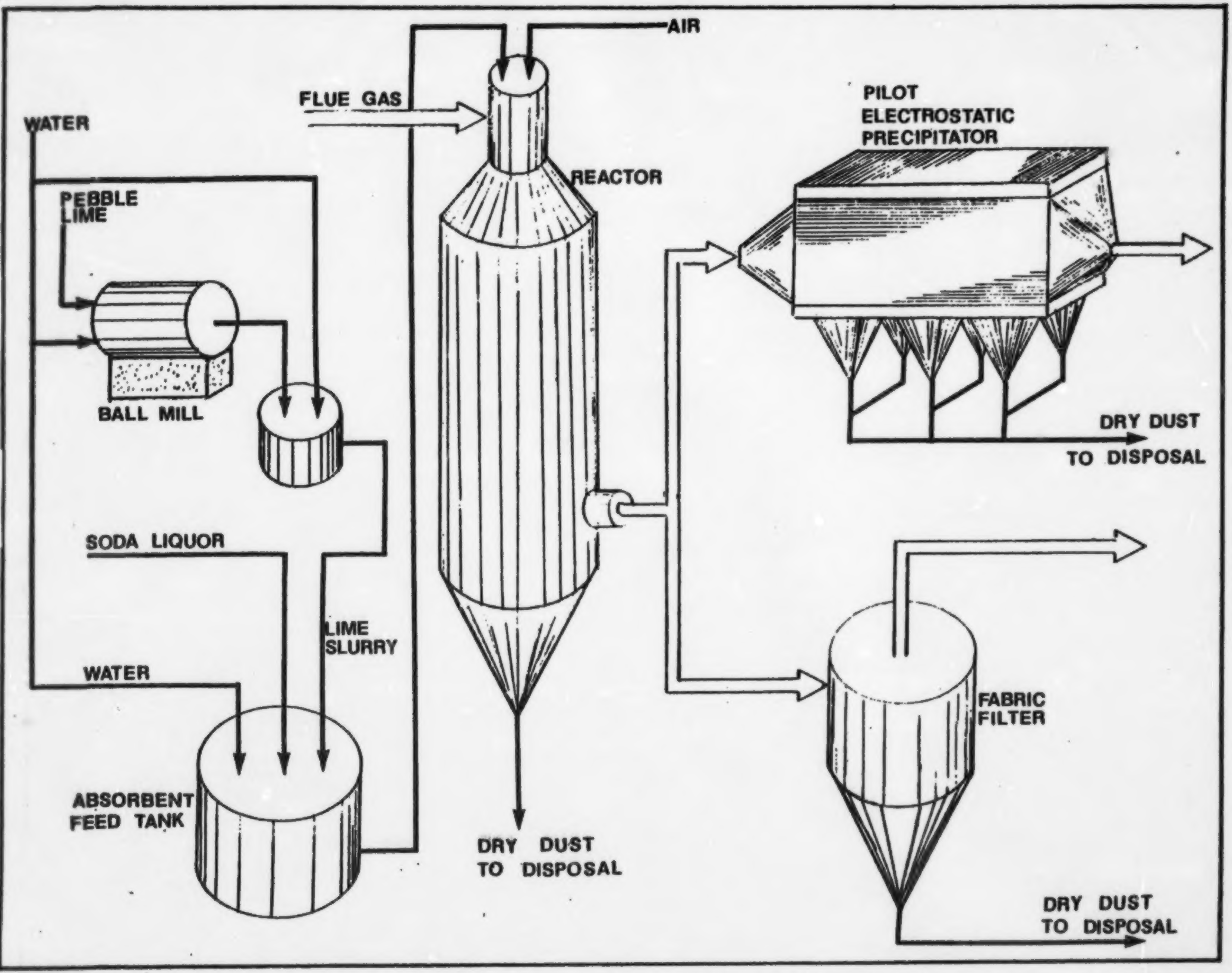
AIR
FLUE GAS
WATER
PEBBLE LIME
REACTOR
PILOT ELECTROSTATIC PRECIPITATOR
BALL MILL
DRY DUST TO DISPOSAL
SODA LIQUOR
LIME SLURRY
WATER
FABRIC FILTER
ABSORBENT FEED TANK
DRY DUST TO DISPOSAL
DRY DUST TO DISPOSAL

FIGURE 4 - JIM BRIDGER DRY SCRUBBING PILOT PLANT PROCESS FLOW DIAGRAM

with a gas discharge nozzle which permits partial gas entry to a pilot EP, and the remainder to a bypass duct. Both towers are carbon steel construction and fully insulated.

For both of the reactors, the incoming flue gas is mixed with the atomized absorbent in the reactor tower inlet disperser. In addition to providing a gas/absorbent mixing zone, the disperser imparts a rotational velocity to the flue gas. This intensifies the mixing with the absorbent and maximizes the effective residence time of the flue gas within the reactor.

The intent behind the reactor design and operation is to assure proper droplet size and gas-liquid contact so that sulfur dioxide mass transfer from the gas phase to the liquid phase is enhanced. This must be accomplished while at the same time assuring a "dry bottom" reactor (i.e., dry, granular dust). The cooled gas (150-175°F) leaves the reactor and then passes into either of two dust collectors.

The pilot electrostatic precipitator, a standard Flakt design and manufacture, is identical to the one used for testing ten years ago to provide the design basis for the existing full-scale precipitators at this site. The significance of this is two-fold.

Firstly, Flakt has developed very comprehensive testing procedures and evaluation methods to assure correct and reliable extrapolation from pilot test data to full-scale design and performance guarantees. Not only has great care been taken in the design of this pilot EP, but also in the design of specialized testing equipment. Secondly, a very unique objective for the Jim Bridger station was the possibility of installing a dry SO_2 absorption reactor upstream of the existing full-scale EP's.

The pilot EP is nominally designed for 3400 ACFM. This means that the FGD pilot system design required partial by-pass of the 15,000 ACFM throughput of the reactor vessel. The pilot EP is of "rigid-frame" design, which represents a "slice" of a full-scale unit, and has the same rapping forces, current density, field configuration and electrode geometry. Each field is equipped with its own high-voltage electrical supply and individual rapping mechanisms for both discharge and collecting electrodes. The discharge electrodes are round spiral type with a diameter of 2.5mm. There are four parallel gas passages through the EP. The collecting electrodes provide an effective collecting area of 825 sq.ft. Dust is collected separately for each of the three fields.

The parallel fabric filter (FF) utilized also represents another innovative concept for the dry SO_2 absorption system. As discussed earlier, this relates to improved SO_2 removal efficiency and improved reagent utilization. The fabric filter is of the high-ratio type with an air-to-cloth ratio of 5-6:1.

There are 16 bags each 6 ft long and made of "Nomex" with on-line cleaning using the "Optipulse" system developed by Flakt. This cleaning method was considered, and, as discusssed later, demonstrated to be the basis for improved performance for the FGD system. The gas for the FF, withdrawn from the outlet ductwork of the reactor was vented to atmosphere after cleaning. Dust was removed through a rotary valve at the bottom of a conical discharge nozzle below the filter.

Two reagents were utilized for the pilot testing: soda liquor and lime. The soda liquor, readily available to the site from local soda ash plants, is already utilized at Jim Bridger as an SO_2 removal agent for a "wet" system. The liquor is delivered by tank truck and stored in an FRP tank with integrally insulated shell and heated by steam to maintain the temperature above 100°F.

Pebble lime, stored in a silo directly above the ball mill, is gravity fed to the mill by means of a screw-feeder. An open circuit wet ball mill was utilized as the slaker with the intent of producing a high quality homogenous lime slurry (approximately 15% solids concentration).

Pumps, piping and fluid control valves are housed in a mechanical enclosure together with the lime slaking equipment. Boiler plant cooling tower blowdown water was used exclusively as the diluent with the only exception being raw (river) water for lime slaking. In addition, an air compressor (rotating-vane type) was utilized, with a maximum discharge pressure of 100 psi.

Finally, a booster fan was used to overcome the pressure losses through the pilot system and to return the gas to the inlet of the full-scale precipitator. A specially designed control house, which contained the data monitoring equipment and electrical controls, was also employed.

Test Program

Testing at the pilot plant was accomplished by nine on-site engineers during the period from March to September of 1980. This field crew was supported by process engineers in Connecticut and Sweden. The initial test plan included eight test blocks which consisted of short-term parametric tests and long-term reliability runs. Later in the program more test blocks were identified and added. Due to the fact that the program was a very ambitious one coupled with a concern for accurate and representative data, the pilot test was predominantly operated 24 hours per day. This assured steady-state operation for the testing.

Operation of the system was dictated, not only by the parametric requirements of each of the test blocks, but also on other factors, such as:

Control Philosophy - due to the interdependent mechanisms which affect the proper operation of this type of system (e.g., the effect of SO_2 inlet concentration, the gas wet bulb temperature, SO_2 removal efficiency, "dry bottom," atomizer feed rate), the design and operation entailed careful consideration for process control sensitivity and responsiveness.

Reactor Tower - continuous, steady-state operation allowed the study of temperature profiles within certain sections of the vessel. Dust flow characteristics were also studied.

Pilot Electrostatic Precipitator - in order to maintain accurate testing, the EP was operated 24 hours per day at constant temperature and flow for each test point. This assured proper conditioning prior to testing. Actual particulate removal efficiencies were based on dust samples at the outlet of the EP and actual gravimetric measurements of dust removed in the precipitator.

Fabric Filter - this filter was operated at a constant pressure drop by means of establishing a proper frequency of on-line cleaning of the bags. Gas flow was maintained constant and adequate time was given for conditioning of the bags to avoid having the filter conditions of the previous test affecting the current test results.

SO_2 Analysis - a critical concern during the testing was to assure the accuracy of the SO_2 readings. The utilization of existing technology with respect to SO_2 analyzers in the relatively new field of "dry scrubbing" has revealed certain problems. The most significant of which is the "baghouse" effect. This phenomenon can very easily occur on the SO_2 analyzing probe located in the duct following the reactor. There is a tendency for dust (containing unreacted alkali) to deposit on the probe screen or filter. This can then react with SO_2 in the sample gas as it is extracted from the flue gas stream. Such a reaction would have the tendency to indicate an artificially higher SO_2 removal efficiency. Corrective maintenance procedures or "in-situ" analyzers are two viable solutions.

Flue gas sampling consisted of pitot tube measurements (to establish and check venturi meter flow measurements), particulate sampling, Orsat analysis, SO_2 analysis (wet chemistry), and wet bulb temperature. Continuous, automatic measurement of gas flows, SO_2 concentration (Du Pont 460), pressure, temperature, liquid flow, pressure drop and opacity (Lear-Seigler RM41) were recorded by a data logger. The capability existed to then input data to an on-site computer, reduce the data, and transmit it to the home office in both Connecticut and Sweden.

An on-site laboratory permitted daily analyses of liquid and solid samples, including sulfite, total oxidizable sulfur, carbonate, sodium and calcium. Flakt's laboratory in Sweden provided additional support for specilized analyses.

Test Results

As shown in Figure 5, the SO_2 removal efficiency improves as the

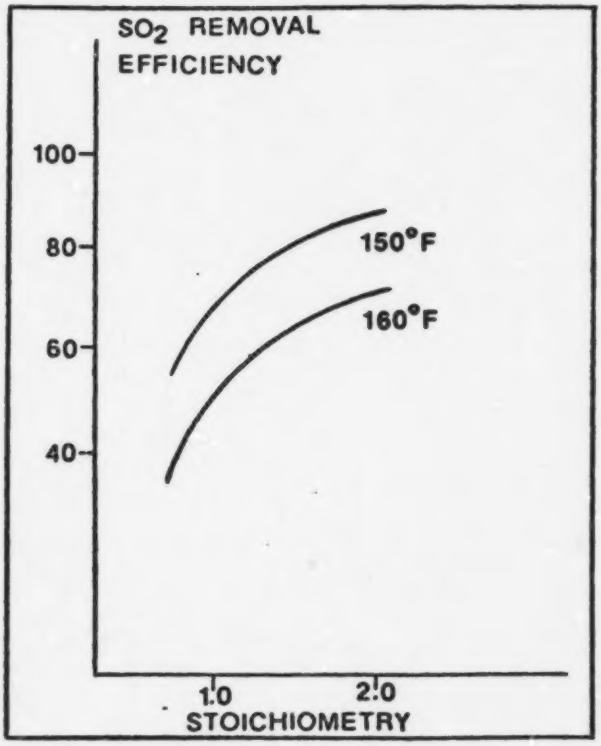

Figure 5: SO_2 removal performance in the reactor only as a function of stoichiometry and reactor temperature using sodium carbonate.

reactor outlet gas temperature approaches the wet bulb bulb temperature. This is consistent with the mass transfer mechanism which occurs between the gas and the liquid droplet; the approach to saturation temperature being a measure of the survival time of the droplet and extent of reaction. The tests showed that the practical limit for operation with the soda liquor (soda ash) absorbent was for the reactor outlet gas to be approximately 20°F above saturation.

In the discussion of results, it must be kept in mind that the objective of high SO_2 removal efficiency must be considered in the light of other operational parameters, the most important of which is "dry bottom." During the start-up and initiation of the pilot program, it was observed that the inlet gas temperature was no greater than $220\text{-}240^{\circ}$F. This low temperature was recognized to be a potentially difficult situation for maintaining dry bottom. It was decided, however, that since this corresponded to both a worst case condition, and a typically low load condition, the tests proceeded. After testing at this condition it was found that dry bottom could be achieved, even at the 20°F approach temperature. Later in the program, higher inlet temperatures were achieved, and as expected, it was significantly easier to achieve dry bottom.

Another important point to consider is the very low alkalinity in the Jim Bridger coal. The benefits of enchanced SO_2 removal efficiencies due to this flyash alkalinity did not exist for the Jim Bridger pilot, so the SO_2 removal performance reflects the effect of the absorbent only.

The other operating parameter which significantly affected SO_2 removal efficiency was the alkali stoichiometric ratio (based on SO_2 in the inlet gas flow). As shown in Figure 5, the greater the alkali input the higher the SO_2 mass transfer. The foregoing results are comparable for both reactors.

It was observed that the SO_2 removal efficiency deteriorated as the inlet SO_2 concentration increased for the same stoichiometric ratio. This could, however, be easily offset by increasing the alkali input further. It is interesting to note that sodium, as expected, was a more reactive alkali than calcium (lime). Unfortunately, it was found that soda liquor and its reaction products were slightly more difficult to dry, as the atomizer droplet size increased. However, by maintaining small enough droplets (i.e., approximately 50μ diameter) it was possible to achieve the efficiencies reported. Further, by increasing the inlet temperature, SO_2 removal efficiency was enhanced.

In the case of the electrostatic precipitator, there was a negligible effect on SO_2 removal efficiency, except that the contact time between reactant and SO_2 in the gas was increased slightly. Particulate removal efficiency, however, was found to be enhanced considerably by the operation of the dry scrubber. Efficiencies of 99.3 to 99.7% were found with correspondingly lower current densities and higher particulate loading at the inlet to the EP. This is attributed to the fact that the migration velocity (W_k) increased due to the lower gas temperature, higher humidity, and higher sodium content of the dust.

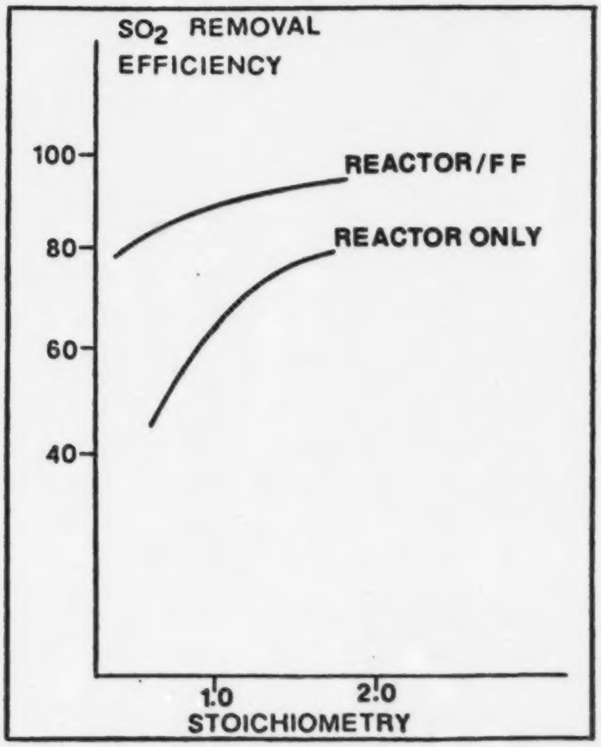

Figure 6: SO_2 removal performance showing the fabric filter contribution with sodium carbonate reagent.

With the study of the reactor performance, it was found that utilization of alkali became lower as the stoichiometry increased. As explained earlier, the use of the fabric filter served to significantly increase utilization in addition to increasing the SO_2 removal efficiency. This is demonstrated in Figure 6 for the sodium case.

This finding becomes even more important in the case of lime. Lime, which is considered to be a more economical alkali, can be used to obtain comparably high SO_2 remvoal efficiency with high utilization. See Figure 7.

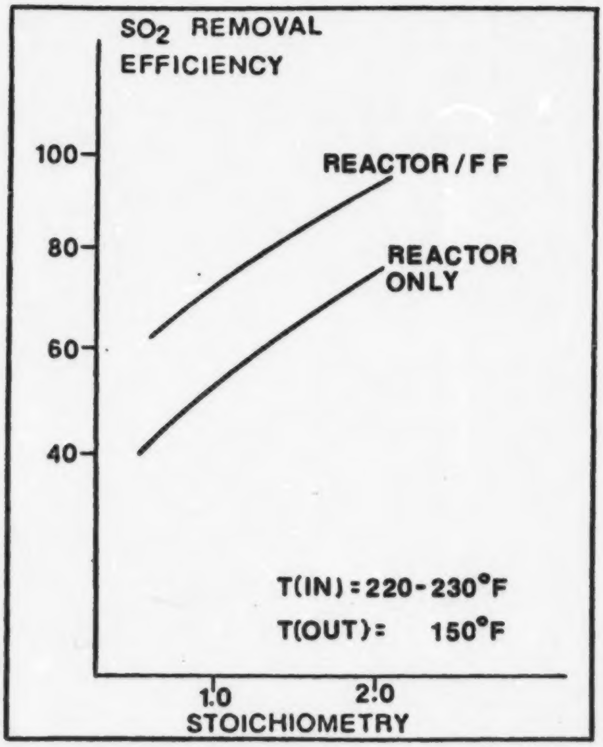

Figure 7: SO_2 removal performance showing the fabric filter contribution with lime reagent.

PERSPECTIVES ON THE DEVELOPMENT OF DRY SCRUBBING--THE COYOTE STORY

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ABSTRACT:

From a historical perspective, development of dry scrubbing systems for SO_2 and particulate control on coal-fired boilers can be traced through a series of steps that culminated in the purchase of the first commercial system in December of 1977. That system will go on line early in 1981 at the new Coyote Station, which is being designed and built by Bechtel Power Corporation for five utility partners (Otter Tail Power Company, Montana-Dakota Utilities Company, Northwestern Public Service Company, Minnkota Power Cooperative, Inc., Minnesota Power & Light Company). The two-stage, dry flue gas cleaning (FGC) system for Coyote is being supplied on a turnkey basis by Rockwell International and Wheelabrator-Frye. This system employs a patented method jointly developed by these two companies.

This paper discusses the history of dry scrubbing as it evolved from the early tests of dry nahcolite injection and the parallel development of spray dryers for flue gas desulfurization. The current status of the Coyote FGC system and the state-of-the-art for dry FGC systems using a variety of alkalies, including lime, are also discussed.

The information provided in the paper will give the utility industry a perspective on how dry scrubbing was developed and how to evaluate the suitability of dry FGC systems for general utility boiler applications.

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PERSPECTIVES ON THE DEVELOPMENT OF DRY SCRUBBING--THE COYOTE STORY

INTRODUCTION

In its broadest context, the term "dry scrubbing" denotes any technique that involves contacting a sulfur-containing flue gas with an aqueous or dry alkaline material and which produces a dry end product. All of the commercial dry scrubbers that have been sold in the United States to date (~4000 MW) are flue gas cleaning (FGC) systems that include (1) a first-stage spray drying device for injecting an aqueous solution or slurry of the alkali into the flue gas, and (2) a second stage, dry particulate collector which removes fly ash and reaction products from the flue gas.

This paper provides a historical perspective on the development and commercialization of two-stage, dry FGC systems. That perspective is based on the events that led to the award of the first contract for a commercial dry FGC system in December of 1977. Subsequent improvements and optimization of dry FGC systems that have occurred in the last 3 years are only discussed in the context of a state-of-the-art summary of the application of these systems on utility boilers.

From the late 1960's through the mid-1970's, the development of two-stage, dry FGC systems proceeded in two separate but parallel paths. The adaptation and development of spray dryers for flue gas desulfurization was begun by Rockwell International in cooperation with Stork-Bowen Engineering Inc. The concept of injecting dry alkalies upstream of a fabric filter was first tested by Southern California Edison (SCE) and further developed by several fabric filter vendors, the most active of which was Wheelabrator-Frye, Inc.

In 1974, the Otter Tail Power Company began to investigate alternatives to wet scrubbers for controlling SO_2 emissions from coal-fired boilers. That investigation initially focused on the technique of dry nahcolite injection into the flue gas upstream of a fabric filter. It became apparent, however, that nahcolite might not be available in sufficient quantities to supply the needs of a large power plant. Bechtel, who was the architect-engineer for Otter Tail's Coyote

Project, recommended an alternate approach, utilizing a spray dryer, which would not be limited to the use of nahcolite. In mid-1977, the Rockwell-Wheelabrator two-stage method was tested on a lignite-fired boiler, using the combination of Rockwell's spray dryer and Wheelabrator's fabric filter. In December of 1977, Bechtel placed the first order for a commercial two-stage, dry FGC system to be installed on the new 410-MW Coyote Station, which is owned by five utility partners (Otter Tail Power Company, Montana-Dakota Utilities Company, Minnkota Power Cooperative, Inc., Northwestern Public Service Company, and Minnesota Power & Light Company).

The two-stage, dry FGC system at Coyote will utilize soda ash as its alkali. This system includes four 46-ft diameter spray dryers, each capable of handling 500,000 acfm of flue gas, and a 38-compartment fabric filter designed to operate with 34 compartments on-line, two compartments undergoing cleaning, and two compartments reserved for maintenance as necessary. The system is presently 95% complete, final startup testing is underway, and commercial operation is scheduled to commence in early 1981.

Since 1977, ten dry FGC systems of various types have been purchased by utilities, and five systems have been purchased for industrial applications. These systems all use lime as their alkali feed, and recent development work has focused on optimization of the lime-based, dry FGC system. Details of ongoing development work by Rockwell, Wheelabrator-Frye, and the other vendors who have entered the market since 1977 can be found in "Survey of Dry SO_2 Control Systems."[1]

FIRST STAGE SPRAY DRYER DEVELOPMENT

The adaptation and development of spray dryers for flue gas desulfurization was begun by Rockwell and Stork-Bowen in 1969. This was not a hardware development program, since the basic hardware had been developed by Stork-Bowen over a 40-year period. The focus of the program was on the physical and chemical aspects of SO_2 removal using a spray dryer. The first tests of a pilot spray dryer on a coal-fired utility boiler were conducted by Rockwell at the Mohave Station of SCE (see Figure 1) in 1972.[2] As a result of this testing, Rockwell made a fixed-price

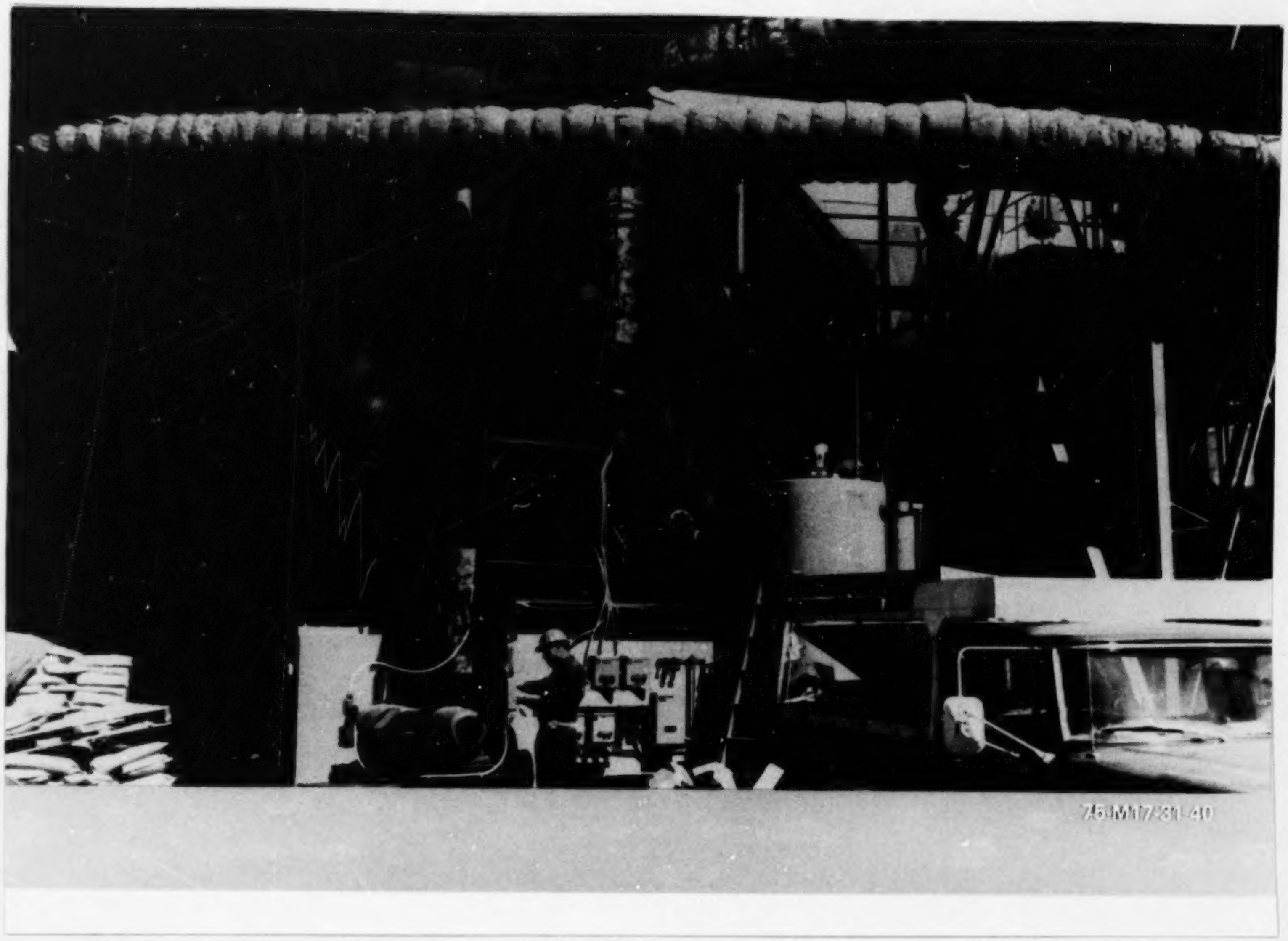

FIGURE 1

MOHAVE SPRAY DRYER TEST FACILITY - 1972

proposal to a western utility in 1973 for a 165-MW system that included two 40-ft diameter spray dryers and an electrostatic precipitator as the particulate collector. In that same time period, the Japanese were also applying spray dryers to oil-fired boilers.[3]

In the time period between 1972 and 1976, Rockwell's spray dryer development work was directed toward the use of soda ash for application in the regenerable Aqueous Carbonate Process (ACP), for which a patent was issued to Rockwell in 1976. Among the objectives of the testing were:

1) to study various atomization techniques including pressure nozzle, two-fluid nozzle, and rotary wheels and select the best for the FGC application,
2) to determine the effect of operating parameters including temperature profiles, SO_2 concentrations, and SO_2 removal efficiences, and
3) to demonstrate the operability of multiple atomizers and study turndown characteristics.

Rockwell has continued its spray dryer development work with extensive pilot and demonstration testing at four utility sites, support testing in a small (250 cfm) dryer, and parametric testing at Stork-Bowen's pilot facility in New Jersey. Much of the recent work has utilized calcium alkalies with the special techniques of gas bypass and/or recycle receiving major attention. To date, Rockwell has conducted about 2000 tests of the application of spray dryers to flue gas cleaning.

SECOND STAGE FABRIC FILTER DEVELOPMENT

The concept of using dry alkali injection upstream of a fabric filter to remove SO_2 and SO_3 from flue gas was first tested by SCE in 1965 at their Alamitos Station. The type of fabric filter utilized in most of the current dry FGC systems was first tested by Wheelabrator-Frye in a series of pilot tests at the Edwardsport Station of Public Service of Indiana between 1967 and 1969.[4] This series of tests included eight different dry alkalies, although only sodium carbonate and sodium bicarbonate were consistently effective. In 1968-1969, Air Preheater Company conducted tests at the Mercer Station of Public Service Electric and Gas Company.[5] These tests showed the strong effect of temperature with best results at flue gas temperatures of about 600°F. The potential advantages of nahcolite were demonstrated by Wheelabrator-Frye at the Nucla Station of the Colorado Ute Electrical Association in 1974.[6] These tests utilized a commercial fabric filter that handled about 65,000 acfm of flue gas.

The most recent work reported was the test series conducted in 1976-1977 by Wheelabrator-Frye and Superior Oil at the Leland Olds Station of Basin Electric in support of Otter Tail's search for a viable dry FGC system. Those tests also involved nahcolite, and the basic results have been reported by Estcourt et al.[7] In that paper, the technique of dry nahcolite injection is compared to a two-stage, dry FGC system, and the latter is shown to be better in terms of potential SO_2 removal efficiency and alkali utilization.

OTTER TAIL POWER AND BECHTEL INVOLVEMENT

When Otter Tail Power first became involved with dry scrubbing, they selected dry nahcolite injection as a promising technique, and small-scale tests were conducted at their Hoot Lake Station by Wheelabrator-Frye in late 1974. The success of those tests led to a pilot demonstration program in 1976, the basic results of which were reported in 1978.[7] When the nahcolite supply problem became apparent in late 1976, V. F. Estcourt, a Bechtel consultant, contacted Rockwell to inquire about the ongoing spray dryer development program. Bechtel, as the A-E on the Coyote Project, was assigned the responsibility of purchasing a workable dry scrubber for the 410-MW Coyote Station, and Mr. Estcourt had been closely following the nahcolite tests. Following his evaluation of the potential of the spray dryer, Rockwell conducted the first tests of the two-stage dry FGC system at Stork-Bowen's New Jersey spray dryer pilot facility in early 1977. The success of that testing led, in turn, to the installation of a pilot spray dryer upstream of the Wheelabrator-Frye pilot fabric filter at Leland Olds (see Figure 2). Demonstration testing commenced in June of 1977, and the test series was completed in August. That program was funded by the

Coyote partners, supervised by Bechtel, and carried out by Rockwell and Wheelabrator-Frye. The first commercial two-stage, dry FGC system was purchased by Bechtel for the Coyote Station in December of 1977. (Rockwell and Wheelabrator-Frye were issued a joint patent on the method utilized in a two-stage, dry FGC system in April of 1980.)

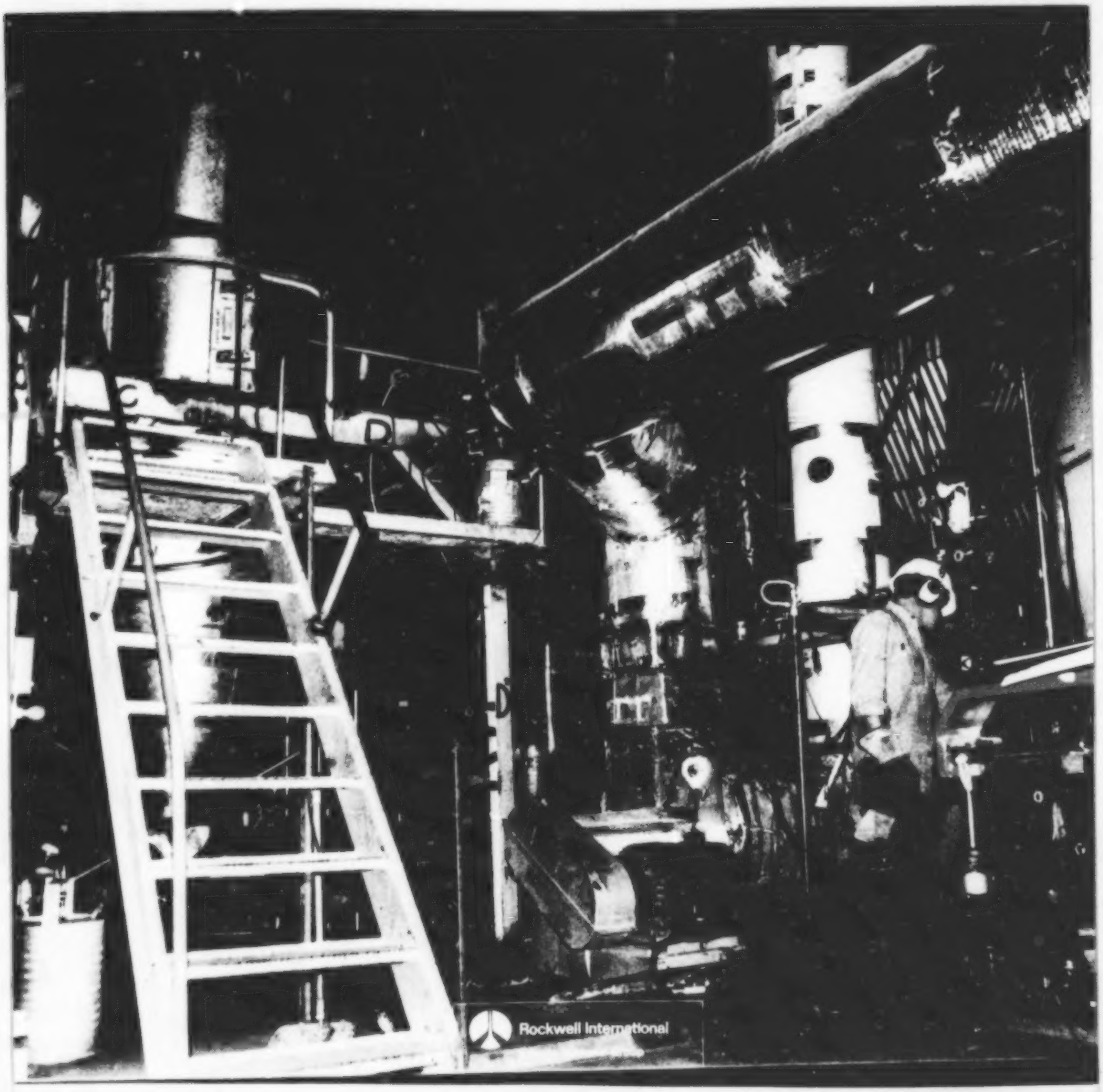

FIGURE 2

LELAND OLDS TWO-STAGE FGC TEST FACILITY - 1977

THE STATUS OF THE COYOTE FGC SYSTEM

The entire Coyote Project is somewhat ahead of schedule, and initial operation of the boiler and FGC system may commence late in 1980. Commercial operation is scheduled for early 1981. Soda ash, stored in bulk as the monohydrate slurry, will be diluted and atomized into the flue gas by twelve, 150-hp centrifugal atomizers. There are three atomizers per 46-ft diameter spray dryer, with each dryer capable of handling 500,000 acfm of flue gas containing up to a nominal maximum of 1400 ppm of SO_2. The Coyote FGC system (see Figure 3) is guaranteed to meet a 1.2-lb $SO_2/10^6$ Btu maximum emission with a guaranteed minimum removal of 70%. Soda ash utilization is expected to approach 100%. Particulate emissions are guaranteed to be less than 0.01 gr/acf, and the clean flue gas will be discharged without reheat (at a temperature of about 180^oF) through a concrete stack with a carbon steel liner. A more complete description of the Coyote FGC system performance and advantages was given by Johnson et al.[8]

Except for the soda ash storage equipment and the atomizing wheel design, the system is suitable for use with any active sodium or calcium alkali. This permits the operating utility (Montana-Dakota) to change alkali with minimal capital cost impact if supplies of an alternate low-cost alkali become available in the future. In fact, the two-stage, dry FGC system can even use nahcolite if its supply situation changes as a result of oil-shale development.

STATE-OF-THE-ART FOR DRY FGC SYSTEMS

Development of FGC systems that use lime as the alkali feed basically commenced in the fall of 1977, although Rockwell had done some preliminary testing as early as 1972. The impetus in the 1977-78 time period was provided by Basin Electric and Stearns-Roger, the A-E for the Antelope Valley Project. Basin Electric was the host utility for the Rockwell-Wheelabrator test program at the Leland Olds Station that resulted in the Coyote award. With Basin's assistance and encouragement, several additional vendors began dry FGC testing including Joy-Niro, Babcock and Wilcox, and Carborundum. Rockwell and Wheelabrator-Frye remained at Leland Olds for an additional year with primary emphasis on lime testing. As a result, dry FGC systems are now commercially available to utilities on a competitive basis with performance that is equal to or better

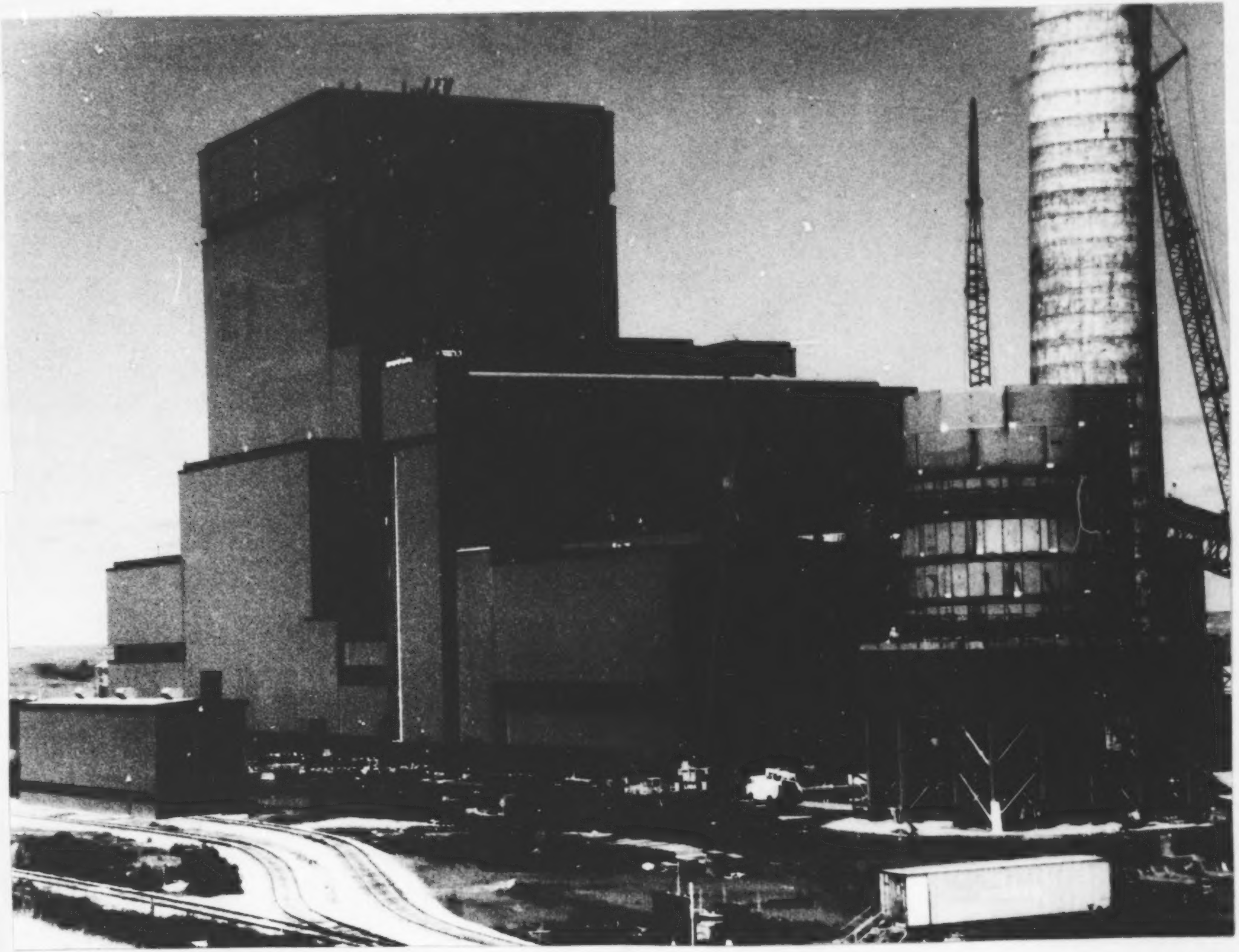

FIGURE 3
THE COYOTE FGC SYSTEM - AUGUST, 1980

than that of wet scrubbers. High efficiency particulate collection is inherent in a dry FGC system, and applicable SO_2 emission standards can be met with 90% or greater lime utilization for SO_2 concentrations from 200 to 2500 ppm.

Currently, there is a general impression that dry FGC is not suitable for high-sulfur coal applications. This impression was reinforced by EPA's 1979 performance standards for utility boilers in which dry scrubbing was somehow tied to the 70% SO_2 removal floor. However, a two-stage, dry FGC system can meet the requisite emission requirements for most utility boilers burning fuel with sulfur content of up to about 4%. The key constraint in the application of dry FGC systems is the relative cost of lime versus the limestone that can be used in a wet scrubber. If, however, the capital cost and other operating cost advantages of the dry systems are carefully evaluated, many utilities burning medium-to-high sulfur-coal may find that dry FGC is competitive with the combination of electrostatic precipitators and wet limestone scrubbers that are needed to meet current emission criteria.

SUMMARY

The unique SO_2 removal concepts and the pioneering development work of Rockwell and Wheelabrator-Frye, the decision by Otter Tail Power to seek an alternative to wet scrubbers, and, most importantly, the encouragement of V. F. Estcourt of Bechtel have resulted in a revolution in flue gas desulfurization technology. There now exists an alternative to wet scrubbers for many utility boiler applications. Each specific FGC application must be evaluated, but in most cases, the two-stage, dry FGC system can be economically applied.

Utilities and architect-engineers all over the country are now involved with the dry FGC system vendors in ongoing development programs and application studies. Since 1977, the list of contributors to the development and acceptance of dry FGC systems is long. For current information on both development and commercial activities, see Blythe, et al.[1]

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KOBELCO
FLUE GAS DESULFURIZATION
PROCESS

Kobe Steel, Ltd.
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INTRODUCTION

Due to the rapid growth of industrial activity after the 1960s, environmental pollution problems in Japan became serious, and especially during the last 10 years, air pollution by sulfur oxides has become an acute social problem.

Therefore, the ambient air quality standard and the emission standard for sulfur oxides have been made rigorous by the Japanese National Environmental Protection Agency and Local Authorities. Under these circumstances, Japanese industries have been forced to develop and install antipollution technology, and KOBE STEEL is no exception. KOBE STEEL is not only one of Japan's leading engineering firms but also a leading steel and machinery manufacturer, and has several emission sources of sulfur oxides in its own works. Therefore, a project team was organized and started research and development of FGD processes in 1972.

As a result, the KOBELCO FLUE GAS DESULFURIZATION PROCESS has been developed as an advanced lime gypsum FGD process.

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2. SHORT HISTORY OF DEVELOPMENT

KOBE STEEL, LTD. has developed a wet scrubbing desulfurization process. During the development stage, our FGD process was applied to 3 kinds of gases as follows.

Test and Pilot Plants

Gas from	Capacity	SO_2 in Gas	Operating Period
Oil fired Boiler	1,000 NM^3/H	500 - 2,000 PPM	Oct. 1972 ～
H_2S Incinerator	12,000 NM^3/H	2,000 - 6,000 PPM	Dec. 1972 ～ Apr. 1973
Iron ore Sinter Pant	50,000 NM^3/H	200 - 400 PPM	Mar. 1973 ～ Jan. 1974

Through the operation of these tests and pilot plants, the design data for a wide range of SO_2 concentration in flue gas was obtained, and according to the data, 4 commercial plants were constructed and the initial plant has been operating for more than 4 years.

In pilot and commercial plants, many troubles have occurred and been solved since the beginning of the development, and simplification of the process has been carried out in commercial plants. As a result, the process has been completed as a simplified FGD process. KOBELCO FGD PROCESS is the only FGD process operating at high concentration of $CaCl_2$ solution in the world and also, only KOBE STEEL, LTD. has operating experience of FGD plants using $CaCl_2$ solution.

Supply list

Customer/Location	On stream date	Capacity	Applied for
Kobe Steel's Amagasaki Works	Jan. 1976	175,000 NM^3/H x 2 trains	Sinter
Kobe Steel's Kobe Works	Feb. 1976	350,000 NM^3/H	Sinter
Nakayama Steel, Ltd.	Jun. 1976	375,000 NM^3/H	Sinter
Kobe Steel's Kakogawa Works	Mar. 1978	1,000,000 NM^3/H	Sinter

Furthermore, this process was exported to Uhde GmbH in West Germany, who is now aggressively working in the European market.

3. FEATURES OF KOBELCO FGD PROCESS

KOBELCO FGD PROCESS is an advanced wet scrubbing desulfurization process in which $CaCl_2$ solution dissolving slaked lime is used as absorption liquid and gypsum is recovered as a by-product. The most notable process feature is that the process is the very best for coal fired boilers' flue gas which contain much HCl as well as SO_2, and the process also has many other features.

3.1 The Optimum Process for Coal Fired Boilers' Flue Gas

The flue gas from a coal fired boiler contains HCl as well as SO_2 and HCl is also absorbed in a SO_2 scrubber and accumulated in the absorption liquid as dissolved $CaCl_2$. In other wet scrubbing processes, to avoid some problems caused by the accumulation, a large amount of waste water must be discharged from the process. But in our process, there are no such problems caused by the accumulation of $CaCl_2$ and no waste water is needed.

(1) No problems caused by $CaCl_2$

$CaCl_2$ in the absorption liquid influences many kinds of design factors of SO_2 scrubbing and also influences the corrosion problems of the process materials. But, KOBELCO FGD PROCESS uses $CaCl_2$ solution originally and has experience of four commercial plants. The initial plant has been operated at high concentration of $CaCl_2$ solution for more than four years. Through the experiences, we can make the best design for SO_2 scrubbing and can also select the best materials for the various conditions of the $CaCl_2$ solution.

(2) No waste water is needed

A balanced amount of $CaCl_2$, which is produced by absorbing HCl in flue gas, must be purged from the system. KOBELCO FGD PROCESS can be operated at 30 - 40 times higher $CaCl_2$ concentration in the absorption liquid than other processes as shown in Fig. 1.

Therefore, there is so little waste water from our process that it can be dried easily and $CaCl_2$ can be recovered as solid $CaCl_2$.

As a result, KOBELCO FGD PROCESS can be applied to the gas from a coal fired boiler as "A CLOSED CIRCUIT FGD PROCESS".

(3) No trouble due to HF

Gases from sinter plants and coal fired boilers contain small amount of HF, and resin lining, rubber lining and brick lining are corroded by HF accumulated in the absorption liquid. However, in $CaCl_2$ solution, HF is precipitated as CaF_2 by the following chemical equilibrium.

$$CaCl_2 + 2HF \rightleftharpoons \underset{\text{(Solid)}}{CaF_2} + 2HCl$$

Therefore, no corrosion problems caused by HF occurs.

3.2 Free from Scale and Slurry Troubles

It is said that scale and slurry troubles are the biggest problems in wet lime-gypsum FGD processes, and the troubles occur in the SO_2 scrubber, pipes and the demister.

However, these troubles were completely solved in our FGD process.

(1) The absorption liquid contains some amount of gypsum solid as seed crystals to prevent scale formation.

Some part of $CaSO_3$ in the absorption liquid is oxidized to $CaSO_4$ by O_2 in the flue gas, so the absorption liquid is kept in a supersaturated concentration of $CaSO_4$. This $CaSO_4$ is crystallized on the inside walls of the equipment and forms hard scale.

But, in the case of existing seed crystals, $CaSO_4$ is crystallized on seed crystals so that no hard scale problem occurs.

(2) In addition to the effect of seed crystals, the solubility and supersaturated range of $CaSO_4$ in $CaCl_2$ solution are much smaller than those in water as shown in Fig. 2.

It is understood by Fig. 2 that $CaSO_4$ can be kept stable in $CaCl_2$ solution.

(3) Viscosity and density of $CaCl_2$ solution are higher than those of water, so the settling velocity of the particles is less than that in water. Therefore, no plugging problem of pipes occurs.

3.3 Low Consumption of Fuel for After-burning

In the case that it is needed to be no white smoke when gas is exhausted from a stack and/or to protect materials of an existing stack from corrosion problems by wet gas, gas from a demister must be heated to 120°C - 140°C by an after-burner or a gas-gas heat exchanger. In KOBELCO FGD PROCESS, temperature of gas from a demister is 5 - 10°C higher than that in other wet scrubbing processes, because of low water vapor pressure of $CaCl_2$ solution.

Therefore, 10 - 20% of fuel for after-burning is saved.

3.4 Recovering Salable Gypsum as a By-product

By-product gypsum produced in the gypsum recovery flow can be used commercially in cement and gypsum board industries. Example quality of by-product gypsum which is produced in our commercial plant is as follows.

Moisture	6 - 10 wt%
$CaSO_4 \cdot 2H_2O$	89%
Cl content	0.02%
Wet Tensile Strength	13 - 17 kg/cm^2

4. PROCESS FLOW

Process flow must be selected in consideration with how to treat a by-product after FGD process. We can provide two kinds of basic process flows.

- Gypsum Recovery Flow

  With an oxidation section and producing gypsum with 10 wt% moisture for commercial use. Refer to Fig. 4.

- Concentrated sludge disposal flow

  With no oxidation section and producing mixture cake of $CaSO_3 \cdot 1/2H_2O$ and $CaSO_4 \cdot 2H_2O$ with 20 - 30 wt% moisture for disposal. Refer to Fig. 5.

A sludge ponding flow and a flow which produces gypsum containing as few impurities as possible can be also provided.

4.1 Gypsum Recovery Flow

The flow shown in Fig. 4 is applied in the case where the flue gas contains few particulates and where it is intended to get gypsum for commercial use.

(1) Cooling and Absorbing

The flue gas is introduced to the lower part of the scrubber and is cooled to 60°C - 70°C and humidified by circulating liquid.

At the same time, some portion of the SO_2 and particulates is removed from the flue gas here.

Finally SO_2 is removed in the upper part of the scrubber by contact with the absorption liquid.

The gas is then sent to stack after passing through the demister.

On the other hand, SO_2 absorbed in the scrubber reacts with slaked lime in the circulating liquid which is a high concentration of $CaCl_2$ solution, and produces $CaSO_3 \cdot 1/2H_2O$.

Some portion of sulfite is oxidized to sulfate by O_2 in the flue gas. CO_2 in the flue gas also reacts with slaked lime as a by-reaction and produces $CaCO_3$.

$$SO_2 + Ca(OH)_2 \rightleftharpoons CaSO_3 \cdot 1/2H_2O + 1/2H_2O ,$$

$$CaSO_3 \cdot 1/2H_2O + 1/2O_2 + 3/2H_2O \longrightarrow CaSO_4 \cdot 2H_2O$$

$$CO_2 + Ca(OH)_2 \rightleftharpoons CaCO_3 + H_2O$$

To treat solid particles of calcium sulfite, sulfate and carbonate, some portion of the circulating liquid is extracted to the thickener. The thickener overflow is recycled to the cycle tank and the thickener underflow is fed to the oxidation section.

(2) Oxidation and Gypsum Treatment

The thickener underflow is mixed with the liquid from the demister and the slurry is acidified to the optimum conditions for oxidation reaction by adding H_2SO_4.

It is then fed to the oxidation tower and the $CaSO_3 \cdot 1/2H_2O$ is oxidized to $CaSO_4 \cdot 2H_2O$ by air from the bottom of the tower. The gypsum slurry from the oxidation tower is concentrated by the thickener and is fed to the basket type centrifuge to be separated into by-product gypsum and the filtrate. The by-product gypsum contains about 10% moisture.

The filtrate is supplied to the scrubber for vaporization into the flue gas when it is cooled adiabatically.

(3) $CaCl_2$ Treatment

HCl in the flue gas is absorbed in the scrubber and produces $CaCl_2$ by a reaction with $Ca(OH)_2$.

In this process flow, $CaCl_2$ is discharged from the system only included with the by-product gypsum and the amount is very small because of the low content of moisture in the by-product gypsum, so $CaCl_2$ absorbed in the scrubber is accumulated in the system.

Therefore, the $CaCl_2$ must be extracted from the system to keep the $CaCl_2$ balance.

Some portion of the thickener overflow is fed to $CaCl_2$ treatment section, in which $CaCl_2$ solution is concentrated by evaporation of water, and is recovered as solid $CaCl_2$.

4.2 Concentrated Sludge Disposal Flow

In the flow, concentrated sludge is produced for disposal, so that there is no oxidation and gypsum treatment section as shown in Fig. 5. And also no $CaCl_2$ treatment section is needed. This is because that the vacuum filter is used to extract the sludge from the system and it contains moisture of 20 - 30 wt%, so waste $CaCl_2$ with the sludge balances that produced in the scrubber.

Fig. 3 shows the balanced concentration of $CaCl_2$ in the absorption liquid for various SO_2 and HCl concentrations in the flue gas.

5. CHEMICAL AND UTILITY CONSUMPTION

Chemical and utility consumption for gypsum recovery flow is as follows:

Plant	Coal Fired Boiler	Iron Ore Sintering
Plant Capacity	500 MW	10,000 t/day
Gas Flow Rate	1,850,000 NM^3/H	1,000,000 NM^3/H
S in Coal	3 %	-
SO_2 Concentration	2,000 PPM	400 PPM
Slaked Lime	12 t/H	1.3 t/H
Sulfuric Acid	1.6 t/H	0.1 t/H
Power	8,300 kW	3,600 kW
Water	85 t/H	40 t/H

Fig. 1 Required Amount of Waste Water From FGD plant for 500 MW Coal Fired Boiler

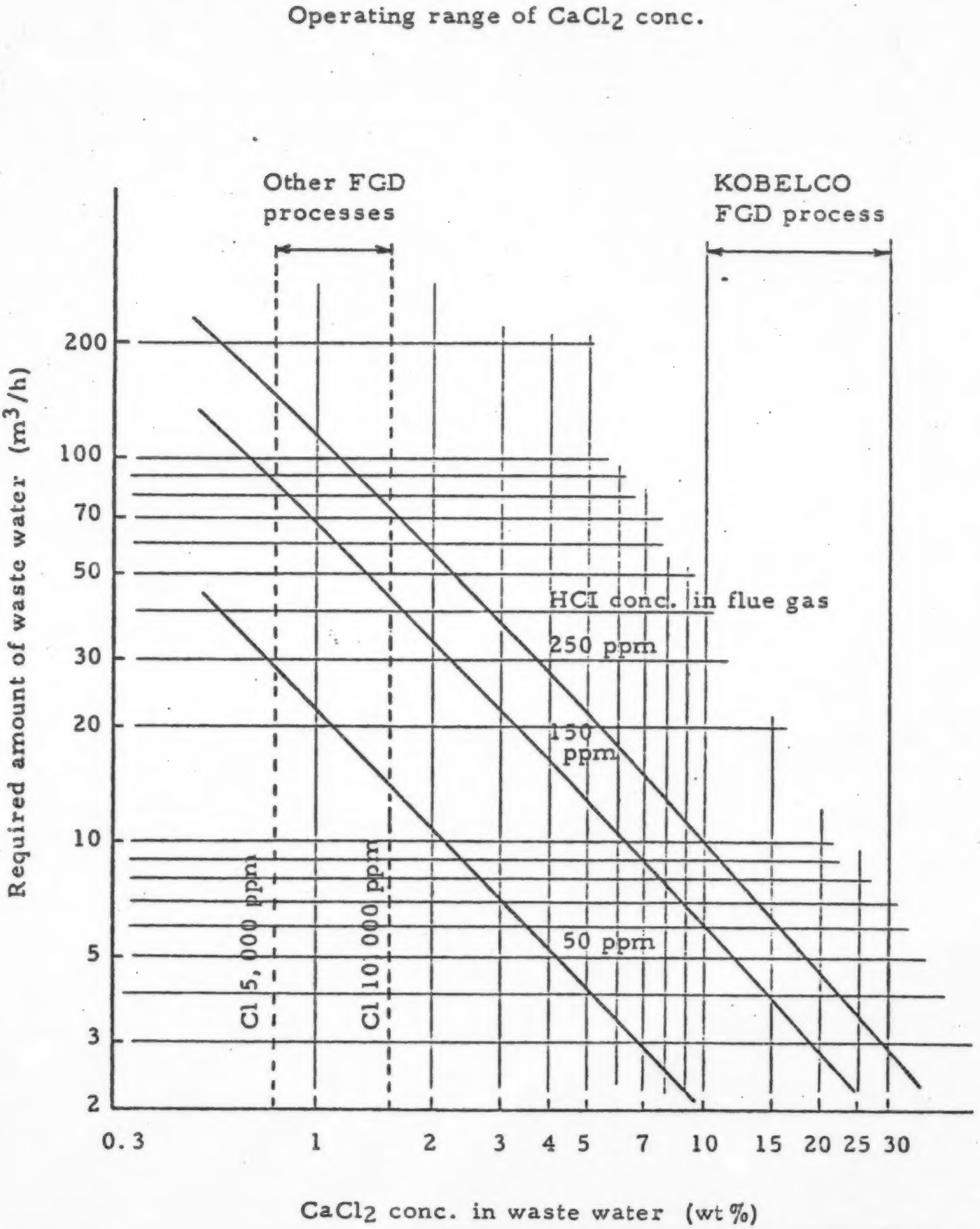


Fig. 2 Supersaturated Concentration of $CaSO_4$ in Water and in 30 wt% $CaCl_2$ Solution

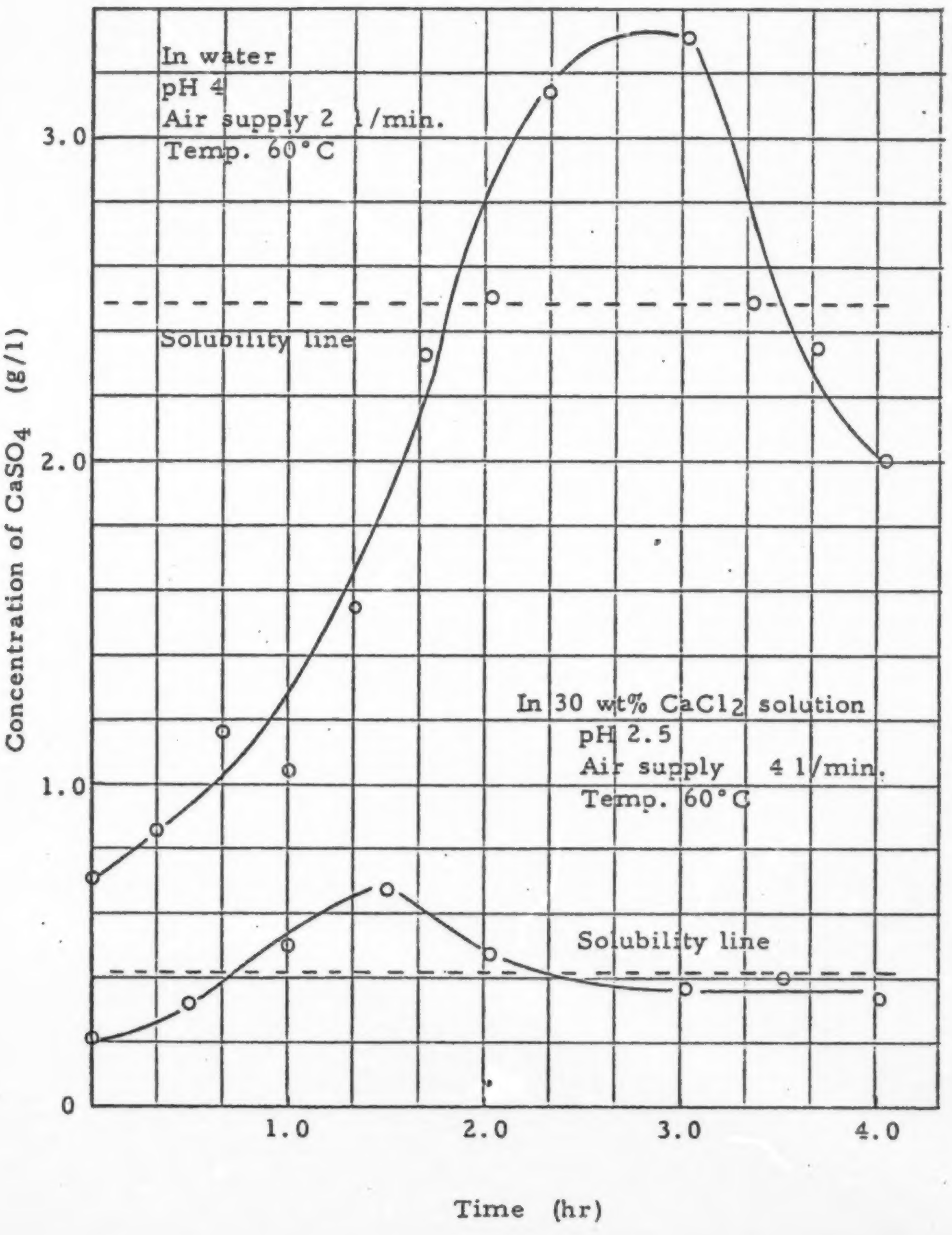


Fig. 3 Balanced $CaCl_2$ Concentration in Absorption Liquid in Concentrated Sludge Disposal Flow

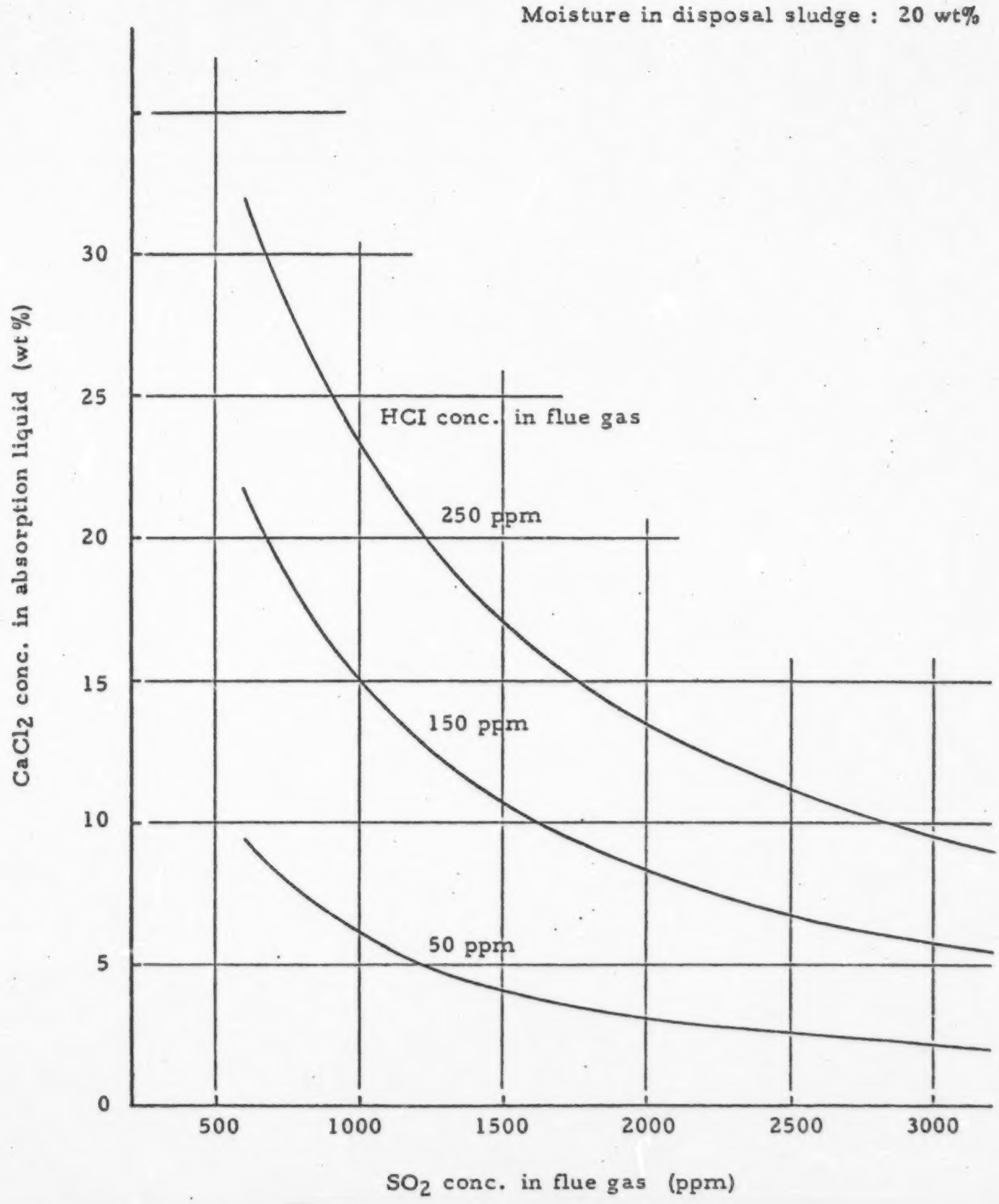


FIG.4 GYPSUM RECOVERY FLOW

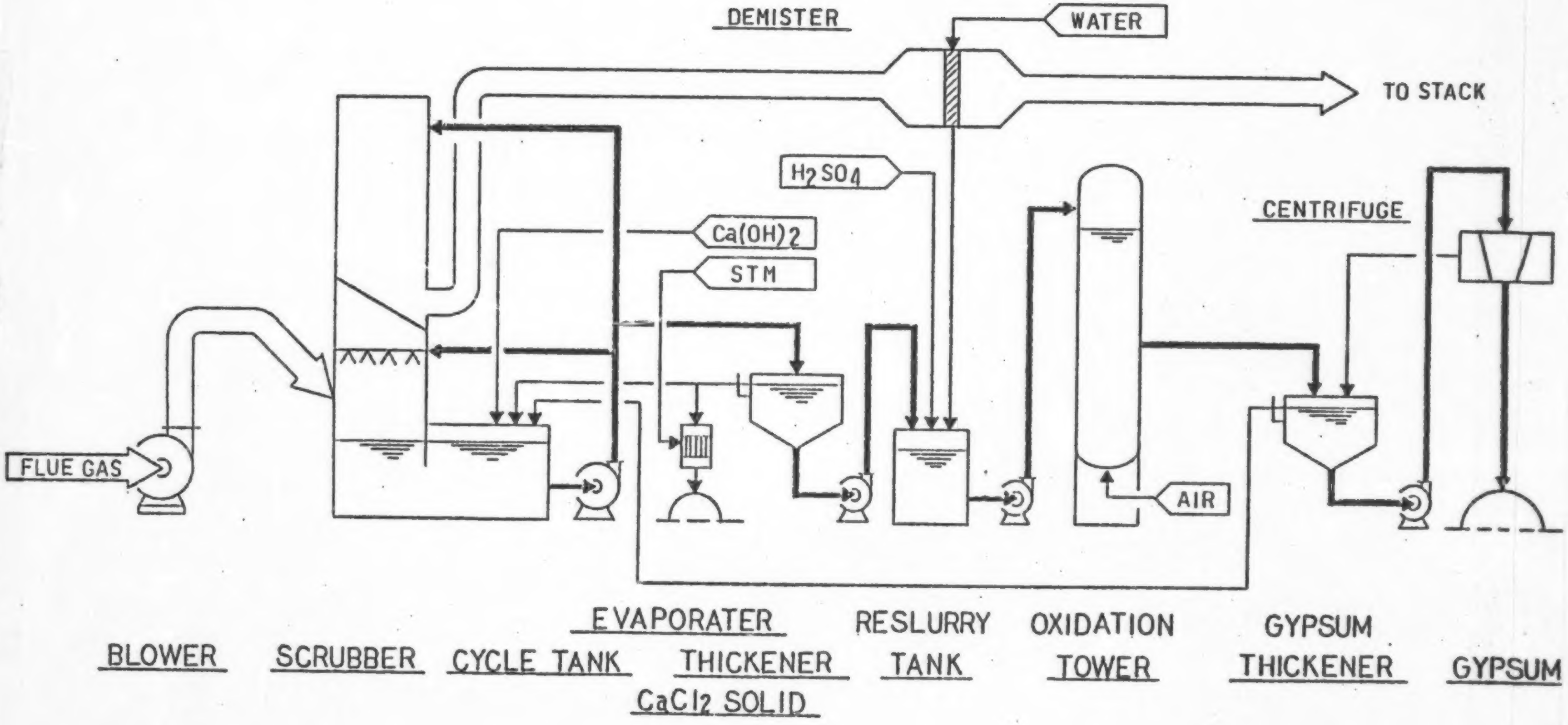
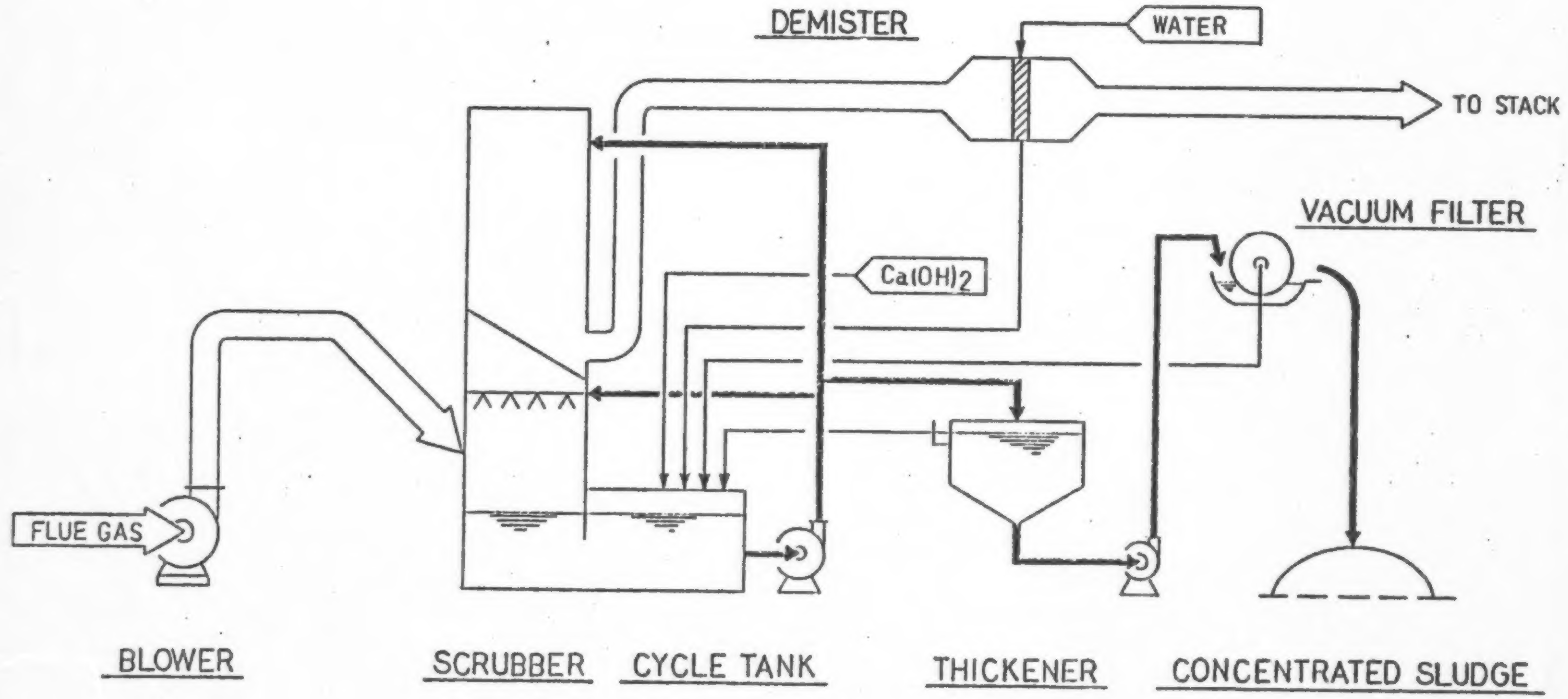

DEMISTER
WATER
TO STACK
H_2SO_4
CENTRIFUGE
$Ca(OH)_2$
STM
FLUE GAS
AIR
BLOWER
SCRUBBER
CYCLE TANK
EVAPORATER
THICKENER
$CaCl_2$ SOLID
RESLURRY
TANK
OXIDATION
TOWER
GYPSUM
THICKENER
GYPSUM

FIG.5 CONCENTRATED SLUDGE DISPOSAL FLOW


DEMISTER
WATER
TO STACK
VACUUM FILTER
Ca(OH)2
FLUE GAS
BLOWER
SCRUBBER
CYCLE TANK
THICKENER
CONCENTRATED SLUDGE

THE RIVERSIDE STATION DRY SCRUBBING SYSTEM

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ABSTRACT

The first full size dry scrubbing system (over 100MW) is scheduled for service in the Fall of 1980 at the Riverside Station of Northern States Power (NSP) in Minneapolis, Minnesota. A description of the plant and dry scrubbing system is provided along with a discussion of the test objectives and reasons for construction of the demonstration plant.

The dry scrubbing system, furnished by Joy Manufacturing Company and Niro Atomizer, Inc., is scheduled to be in service before any other utility dry scrubbing plant and the results will be of importance to all who are considering future flue gas desulfurization systems.

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THE RIVERSIDE STATION DRY SCRUBBING SYSTEM

BACKGROUND

Since it was first publicized in 1977, the concept of flue gas desulfurization by spray dryer absorption with lime has generated a remarkable degree of acceptance by the electric utility industry. In the United States the first pilot plants were built in 1976 and 1977 and economic utilization of lime as a reagent was first demonstrated in 1978. Within about two years of this first demonstration, there have been ten full size electric utility dry scrubbing systems contracted for, along with four industrial plants. Additional dry scrubbing systems are currently in negotiation and being specified. Table 1 summarizes those contracts which have been awarded to date. At least 20 pilot plants have been built by several manufacturers to test and demonstrate the dry scrubbing process. (1)

The total dollar value of dry scrubbing systems which have been awarded to date or which are currently being negotiated or specified probably exceeds a half billion dollars. The value of the power plants to be served by these systems is believed to be between four and five billion dollars. With this sort of an investment, there is naturally a great deal of interest by the electric utility industry in the performance of a full-sized installation.

The concept of dry scrubbing for flue gas desulfurization has been well documented. Meyler (2) has identified 15 technical papers and reports in the bibliography of a recently published paper. Therefore, it will suffice to summarize that the dry scrubbing process consists of introducing hot, untreated flue gas into a spray dryer absorber. Here, the gas contacts a finely atomized spray of alkaline slurry (usually lime) so that sulfur dioxide is absorbed into the spray droplets as water is simultaneously evaporated. The flue gas next passes into particulate collection equipment (a baghouse or precipitator) where flyash and reaction products are removed from the gas stream. The cleaned flue gas then flows through induced draft fans and out the stack.

The two basic advantages of dry scrubbing as compared to more conventional wet scrubbing are simplicity and economy. Burnett et al. (3) prepared a preliminary economic analysis of two flue gas desulfurization processes, one dry and one wet, for a new 500-MW power plant burning western coal. Results of the analysis show that the capital investment costs for the dry scrubbing system are $132/kW compared to $186/kW for a typical electrostatic precipitator - wet limestone scrubber combination. Levelized annual revenue requirements are 8.55 mills/kWh for the dry FGD process and 11.71 mills/kWh for the wet system.

Joy Manufacturing Company and Niro Atomizer, Inc. are leaders in the field of dry scrubbing. At about the time of the 1979 FGD symposium in Las Vegas, it became apparent to the two companies that the concept of dry scrubbing was rapidly gaining acceptance in the utility and industrial market place. However, many utility representatives were still expressing a desire to see a full-size system in service before purchasing one themselves.

At that time it was estimated that the potential market for dry scrubbing systems in the United States through 1982 could exceed three billion dollars; however, the first lime spray dryer system was not scheduled for start up until 1982. Many utilities could potentially make a decision to install wet scrubbing systems while waiting for dry scrubbing to be commercially demonstrated.

Joy and Niro decided to try to find a site for a system large enough to be demonstratably acceptable to the utility industry. A minimum size requirement was considered to be 100 MW. An existing plant appeared to be a necessity in order to expedite construction. It was, of course, hoped that a host utility might be found which would be willing to share in the cost of such a demonstration plant.

NSP and their engineers, Black & Veatch, had prepared and issued specifications for a dry scrubbing system for the proposed 800-MW generating unit number 3, at their Sherburne County Generating Plant near Becker, Minnesota. One of the requirements of the specification was that the bidder should provide a pilot plant to test the flue gas from the Absoluka Mine coal being burned at that plant.

Joy and Niro countered with an offer to build a 100-MW demonstration module to operate in parallel with the existing wet scrubbers on Sherburne Unit #2. This received initial favorable consideration by NSP. However, due to a delay in the construction of Unit #3, the initial bids were rejected by NSP. They in turn suggested an alternate location at their Riverside Station which, for reasons to be discussed below, offered advantages to all parties. In September of 1979 a contract was signed to build the demonstration plant at the Riverside Station. Black & Veatch was selected as the engineer for NSP.

THE RIVERSIDE DEMONSTRATION PLANT

Riverside Station is the oldest fossil-fueled generating station in the NSP system. Units 6 and 7 were built in 1949 and 1953 and include two pulverized coal-fired Babcock and Wilcox boilers, both with steam generating capabilities of about 65 MW (net). Each boiler is followed by two parallel Research-Cottrell wire-and-weight type electrostatic precipitators. An induced draft fan follows each precipitator and both boilers have their own stack.

The plant originally burned Illinois coal with a sulfur content of 3% or more. When environmental requirements necessitated a reduction in sulfur oxide emissions, NSP made a decision to burn low sulfur western coal in Units 6 & 7 at Riverside. This reduced the sulfur oxide emissions, but the low sulfur coal caused problems in the Riverside precipitators because of increased flyash resistivity. NSP now had a particulate problem. This was alleviated by operating the boilers at reduced load and also by mixing some high sulfur coke with the low sulfur coal in each boiler so that sulfur oxide requirements were not exceeded but resistivity was improved.

Still, it was necessary to operate the boilers at reduced load. Later, when an accident occurred damaging the turbine generator at Riverside Unit #7, it was decided to remove the turbine from service, and the boilers now provide common steam to Riverside Unit #6 with some excess steam used elsewhere in the plant.

Installation of a 100-MW dry scrubbing system at Riverside (approximately 500,000 acfm) would permit use of any coal in the Riverside boilers to generate more than their current steam capabilities. A baghouse, incorporated in the dry scrubbing system, would be sized for about 420,000 acfm, due to the reduced gas volume caused by the lower spray dryer exit temperature. NSP wanted a baghouse sized to handle the full potential gas volume from the two boilers (approximately 640,000 acfm with spray dryer out of service) and agreed to pay for the incremental cost of enlarging the baghouse from eight to twelve compartments with provisions to add two additional compartments if required in the future.

Therefore the installation of the dry scrubbing demonstration unit at Riverside provides NSP with the potential for increased generating capacity while solving a troublesome particulate problem and gaining direct experience in the operation and maintenance of a dry scrubbing system.

Riverside provides an ideal location for a demonstration plant. Low, medium and high sulfur fuels can be burned in the plant, and testing can be done on each, individually or in combination. Since the generating units are relatively small, a reasonable degree of flexibility is available for testing purposes. A schematic diagram of the system is shown in Figure 1.

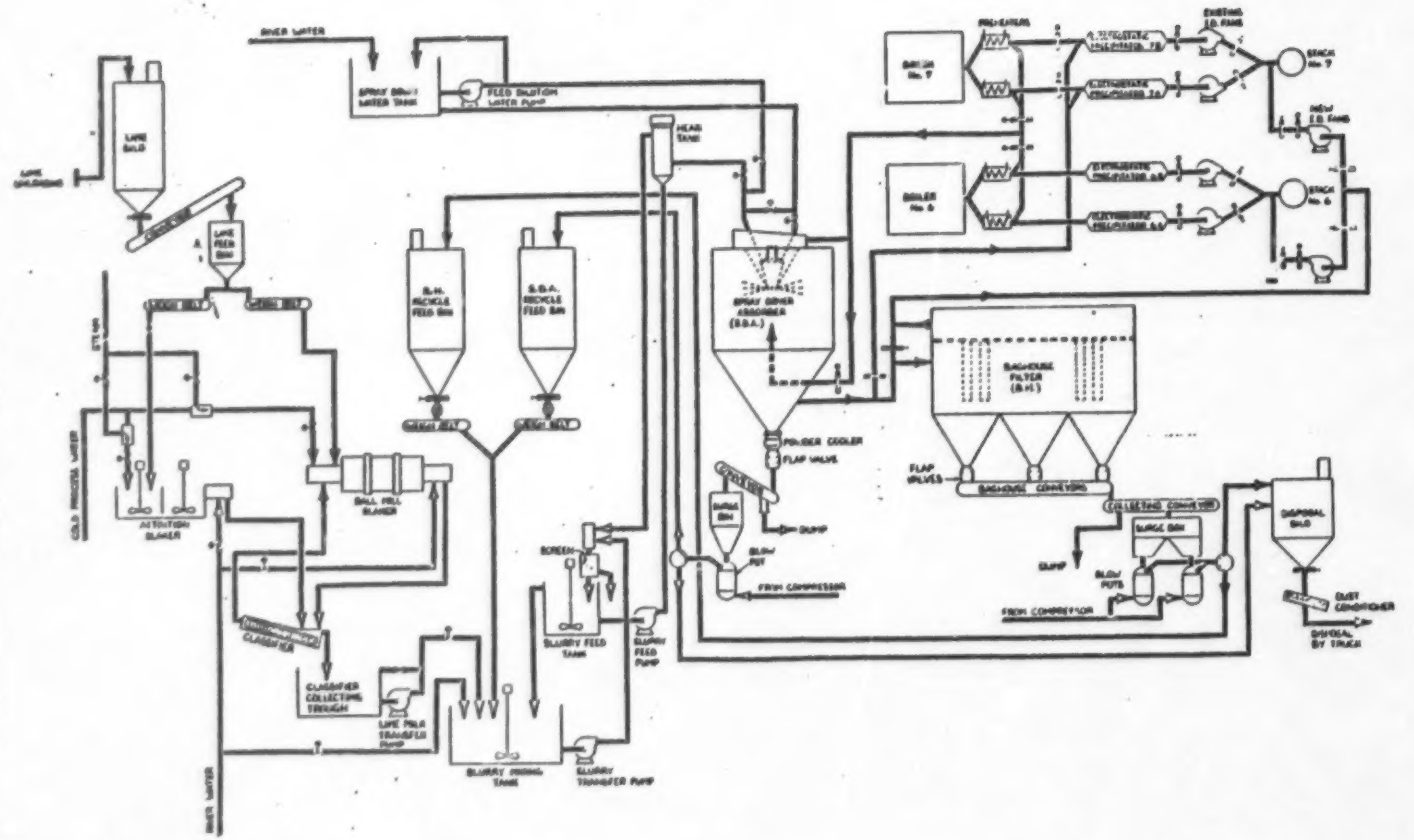
Having existing precipitators in the plant along with a new baghouse provides the opportunity to test dry scrubbing with both. Earlier pilot tests indicated better results in a baghouse; however, testing with full sized equipment will provide more meaningful data. Inserting a spray dryer and ductwork ahead of existing precipitators and ID fans will add additional draft loss to the system and therefore full load operation cannot be accomplished with precipitators. However, useful performance results can be obtained operating the precipitators and baghouse at comparable reduced loads.

The spray dryer absorber at Riverside is identical to those being furnished by Niro Atomizer for several other dry scrubbing systems currently in design and construction. It is larger than any other spray dryer operating in the world. The demonstration unit will provide Joy and Niro with the opportunity to verify the scale-up concepts used in its design. The maximum potential flue gas volume from the boilers will provide an opportunity to observe performance of this unit at approximately 25% above design conditions.

The baghouse is the first of a new structural design concept offered by Joy. Fabrication and construction costs are being closely monitored to provide accurate input data for future estimates. The baghouse will be the first in a power plant to be operated consistently at temperatures as low as 10°C above the adiabatic saturation point. The cold Minnesota climate is expected to help demonstrate that the unit can be operated without problems under cold weather conditions. Several different filter fabrics will be tested at Riverside.

Various process options and alternate equipment designs will be installed and tested which will be helpful in offering more competetive systems in the future.

Control for the system will be fully computerized, and the demonstration plant provides an opportunity to prove out various concepts in hardware and software design.


RIVER WATER
SPRAY DRYER WATER TANK
FEED DILUTION WATER PUMP
LIME SILO
LIME UNLOADING
LIME FEED BIN
STEAM
COLD PROCESS WATER
ATTRITION SLAKER
BALL MILL SLAKER
CLASSIFIER
CLASSIFIER COLLECTING TRUGH
LIME MILK TRANSFER PUMP
RIVER WATER
B.H. RECYCLE FEED BIN
S.D.A. RECYCLE FEED BIN
SLURRY MIXING TANK
SLURRY TRANSFER PUMP
SCREEN
SLURRY FEED TANK
SLURRY FEED PUMP
HEAD TANK
SPRAY DRYER ABSORBER (S.D.A.)
POWDER COOLER
FLAP VALVE
SURGE BIN
BLOW POT
DUMP
FROM COMPRESSOR
BOILER No. 7
BOILER No. 6
PREHEATERS
ELECTROSTATIC PRECIPITATOR 7B
ELECTROSTATIC PRECIPITATOR 7A
ELECTROSTATIC PRECIPITATOR 6B
ELECTROSTATIC PRECIPITATOR 6A
EXISTING I.D. FANS
STACK No. 7
NEW I.D. FANS
STACK No. 6
BAGHOUSE FILTER (B.H.)
FLAP VALVES
BAGHOUSE CONVEYORS
COLLECTING CONVEYOR
SURGE BIN
DUMP
BLOW POTS
FROM COMPRESSOR
DISPOSAL SILO
DUST CONDITIONER
DISPOSAL BY TRUCK

FLOW SHEET -
RIVERSIDE DRY SCRUBBING SYSTEM

FIGURE 1

Equipment furnished by Joy and Niro includes connecting ductwork, spray dryer absorber, baghouse, ash and waste product handling to the disposal bin, lime unloading, storage and conveying equipment, slaking and process equipment, support structures, motor control centers, instrumentation and computer control. Additionally, Joy and Niro are providing installation of the above equipment as well as supervision of the initial operation, testing and startup of equipment.

The test objective is to remove 90% of the sulfur dioxide from the flue gas and reduce particulate emission to an outlet loading of .03 lb/MM Btu.

TEST PLAN

The unit will first operate with precipitator dust collection in the Fall of 1980, and the baghouse is scheduled to come on line by January, 1981. Some of the initial tests will include operating the equipment at various loads with different spray dryer outlet temperatures, slurry concentrations, atomizer wheel speeds and SO_2 collection efficiencies.

It is expected that these short term tests will be completed early in 1981 and a performance demonstration can be started about that time. An investigation program is then planned consisting of the following elements:

- Demonstration of operation with at least three different fuels at specific SO_2 removal levels for each fuel, operating one week or more at each fuel/efficiency combination to assure system stabilization.
- Analysis for disposal properties of representative waste product from selected runs.
- Documentation and demonstration of startup, shutdown, and load-following procedures.
- Collection of operating and maintenance cost data.
- Characterization of particulate matter emitted from the system.
- Development of availability, reliability, and operability data on the system.

In demonstrating the system capability with respect to these tests, the following measurements are expected to be made and reported:

- Emission mass concentration and particle size of particulate matter.
- SO_2 inlet and outlet concentrations.
- Boiler load and firing conditions.

- Coal analysis and flow rate.
- Lime analysis and utilization.
- Gas volume flows and temperatures.
- Chemical and mechanical analysis of waste products.
- Number of operator manhours.
- Number of maintenance manhours.
- Spare parts usage.
- Hours down-time vs. hours operating.
- Behavior of corrosion test coupons.
- Utilities requirements.
- General descriptive information

CONCLUSION

The Riverside Dry Scrubbing Demonstration Plant represents a substantial investment by private industry. However, the benefits to the investors and to the electric utility industry in general should prove to be well worth the cost.

This installation will be in operation this fall. It is full-scale and has testing facilities in excess of normal commercial requirements. The on-site computer will provide data storage, reduction and display, and the station load and fuel schedules are flexible. All of these features provide a highly attractive opportunity for realistic assessment of the performance of dry scrubbing.

TABLE 1.

COMMERCIAL DRY SCRUBBING SYSTEMS - 7/31/80

OWNER	PLANT	SIZE	ENGINEER (Utility Pow. Plant)	SUPPLIER
Otter Tail Power Company	Coyote #1	410 MW	Bechtel	RI-WFI (1)
Basin Electric Power Cooperative	Antelope Valley #1	430 MW	Stearns-Roger	Joy/Niro (2)
Basin Electric Power Cooperative	Laramie River #3	500 MW	Burns & McDonnell	B & W (3)
Northern States Power Company	Riverside #6 & 7	130 MW	Black & Veatch	Joy/Niro
Tucson Electric Company	Springerville #1 & 2	2-350 MW	Bechtel	Joy/Niro
United Power Association	Stanton	60 MW	Black & Veatch	R-C (4)
Colorado-Ute Electric Association	Craig #3	450 MW	Stanley Consultants	B & W
Platte River Power Authority	Rawhide #1	250 MW	Black & Veatch	Joy/Niro
Sunflower Electric Cooperative	Holcomb #1	310 MW	United Engineers	Joy/Niro
Strathmore Paper Company	Woronoco, Mass.	40,000 acfm		Mikropul Corp.
Celanese Fibers Corporation	Cumberland, Md.	65,000 acfm		RI-WFI
University of Minnesota	Minneapolis, Minn.	120,000 acfm		Carborundum (5)
Calgon Corporation	Big Sandy, Ky.	51,000 acfm		Niro/Joy

(1) Rockwell International and Wheelabrator Frye, Inc.
(2) Joy Manufacturing Company & Niro Atomizer, Inc.
(3) Babcock & Wilcox
(4) Research-Cottrell
(5) Carborundum Environmental Systems Division of Kennecott Copper Company

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1. G. M. Blythe, J. C. Dickerman and M. E. Kelly - Survey of Dry SO_2 Control Systems. U.S. E.P.A. Publication 600/7-80-030, February 1980.

2. Meyler, J. A. - Dry Flue Gas Scrubbing - A Technique For The 1980's (Presented at the 1980 Joint Power Generation Conference. Phoenix, Arizona. September, 1980).

3. T. A. Burnett and W. E. O'Brien - Preliminary Economic Analysis of a Lime Spray Dryer FGD System. U.S. E.P.A. Publication No. 600/7-80-050, March 1980.

EVALUATION OF GYPSUM WASTE DISPOSAL BY STACKING

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ABSTRACT

Forced oxidation flue gas desulfurization (FGD) scrubbers can produce significant quantities of waste gypsum, which if not utilized, requires safe and economical disposal. Gypsum is also a waste product of the phosphate fertilizer industry which has successfully utilized stacking methods of waste disposal for more than 20 years. Results from geotechnical laboratory testing of Chiyoda Thoroughbred 121 (CT-121) FGD gypsum are presented. These results indicate FGD gypsum has settling, dewatering, and structural characteristics similar to and, in some instances, more favorable than phosphate gypsum, making stacking methods of waste disposal a possible option for disposing of FGD gypsum. The construction and nine-month operation of a one-half acre, 12-foot high prototype CT-121 FGD gypsum stack at Plant Scholz is also discussed. The success of this installation further confirms the feasibility of utilizing stacking for disposal of FGD gypsum.

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EVALUATION OF FGD GYPSUM WASTE DISPOSAL BY STACKING

INTRODUCTION

Currently, and for the past several years, three alternatives have been available to electric utilities to meet federal new source performance standards (NSPS) for coal-fired electric generating stations: (1) burning a coal whose combustion products complied with these regulations, (2) cleaning the combustion gases from noncomplying coal, or (3) cleaning noncomplying coal prior to combustion. The Clean Air Act Amendments of 1977 (Public Law 95-95) have effectively eliminated the first option and placed further constraints on the latter two. Because of this, the development of alternative technologies for postcombustion cleaning of flue gases and precombustion cleaning of coal has taken on a new importance.

Since 1972, Southern Company Services, Inc. (SCS) has been extensively evaluating postcombustion cleaning of flue gas at the Scholz Electric Generating Station (Scholz) of Gulf Power Company in Sneads, Florida (1). As a continuing part of the evaluation program at Scholz, Chiyoda International Corporation installed and operated a 20 MW prototype of their Thoroughbred 121 (CT-121) forced-oxidation direct limestone flue gas desulfurization system (2). The evaluation of the CT-121 system, sponsored by the Southern Company and the Electric Power Research Institute (EPRI), included an overall process evaluation by Radian Corporation (RP536-4) and an evaluation of the feasibility of utilizing stacking for waste disposal of CT-121 FGD gypsum by Ardaman & Associates, Inc. (RP536-3).

Project Purpose and Objective

The application of forced oxidation for converting calcium sulfite sludge to calcium sulfate (gypsum) in flue gas desulfurization (FGD) sludge processing and disposal has recently received increased attention. The mineralogy, crystal geometry, and particle size of gypsum typically provided settling, dewatering, and structural characteristics which allow easier and more efficient methods of waste disposal than with calcium sulfite sludges.

Gypsum is also a waste by-product of the phosphate fertilizer industry. The greatest United States concentration of phosphate mining and processing exists in Florida. Over 21 million tons (1.9×10^{6} kg) of waste by-product gypsum is produced annually. Stacking methods of waste disposal have been economically utilized by the phosphate industry in Florida for gypsum disposal for more than 20 years. These gypsum stacks are typically large (50 to 300 acres), structurally stable stockpiles reaching heights of 150 feet (45 m). A typical phosphate industry gypsum stack located near Bartow, Florida is shown in Figure 1.

Although gypsum stacking has been successfully utilized by the phosphate fertilizer industry for waste disposal, no experience exists on the stacking behavior and engineering characteristics of waste gypsum produced by FGD scrubbers. An objective of the CT-121 process evaluation, therefore, was to study the geotechnical and environmental feasibility of utilizing stacking methods of waste disposal for CT-121 FGD gypsum. To achieve this objective, a prototype gypsum stack was constructed and operated over the nine-month CT-121 process test period at Plant Scholz from October 1978 to June 1979.

The results obtained from the Plant Scholz study are presented herein and address three items: (1) an engineering evaluation of the geotechnical properties of CT-121 FGD gypsum, (2) a summary of stacking operations and stack performance at Plant Scholz, and (3) an overall appraisal of the feasibility of utilizing stacking for disposal of FGD gypsum.

Background of FGD By-Product Gypsum

The potential for utilizing stacking methods of waste disposal for FGD gypsum was initially recognized during operation of the Chiyoda Thoroughbred 101 (CT-101) scrubber at Plant Scholz (1). The CT-101 FGD system produced an essentially pure gypsum which could be dewatered by centrifuge to 80 to 88 percent solids. The dewatered CT-101 FGD gypsum was hauled by dump truck to a lined, above-ground storage pond for disposal. Although the applicability of gypsum stacking, as carried out by the phosphate industry, was recognized as a possible method of waste disposal, no stacking experiments were included in the CT-101 process evaluations.

Prior to the CT-121 process installation at Scholz, the engineering characteristics of CT-121 FGD gypsum from a Chiyoda pilot plant in Japan were evaluated (3). Laboratory tests were conducted to determine the engineering properties

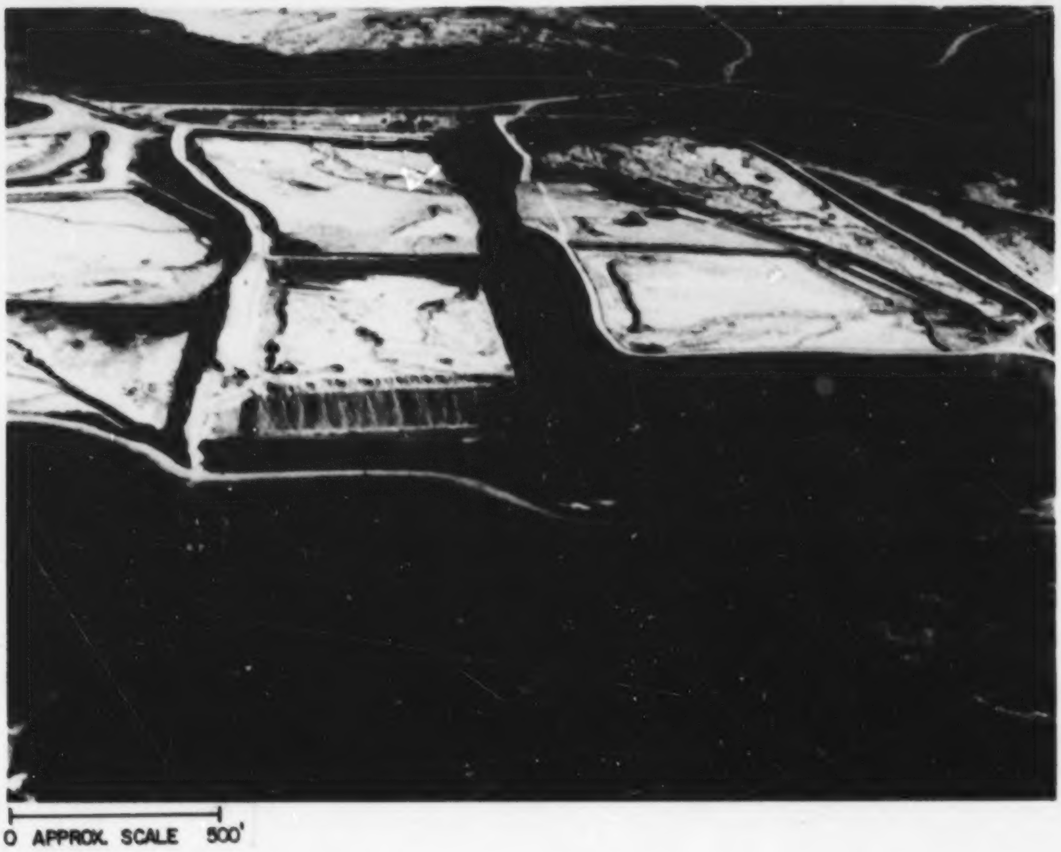
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Figure 1. Phosphate Fertilizer Plant Gypsum Stack Near Bartow, Florida.

of CT-121 FGD gypsum relevant to the stacking method of disposal, and for comparison to the properties of typical phosphate gypsums, which have been successfully stacked. This study indicated that the pilot plant CT-121 FGD gypsum had engineering properties similar to phosphate gypsums and that a prototype stack could probably be constructed without difficulty. Therefore, during the subsequent installation and operation of the CT-121 Scrubber at Plant Scholz, the opportunity was offered to construct and study the performance of a prototype FGD gypsum stack.

CT-121 PROCESS DESCRIPTION

The CT-121 process is similar to that of conventional limestone scrubbing processes, but different in that SO_2 is completely oxidized to calcium sulfate (gypsum) in the absorber. The oxidation is so complete that only trace amounts of calcium sulfite can be found. A schematic process flow diagram of the CT-121 process is shown in Figure 2 (2). Flue gas from the plant, after quenching, is introduced directly into the Jet Bubbling Reactor, where it is then sparged into the absorbent through an array of vertical spargers, generating a jet bubbling layer. Sulfur dioxide (SO_2) is absorbed in the jet bubbling layer producing calcium sulfite which is oxidized completely to calcium sulfate. The cleaned flue gas then flows from the reactor, through a mist eliminator, and out the stack.

Limestone slurry is pumped directly to the Jet Bubbling Reactor to precipitate sulfates as gypsum. The crystallized gypsum is discharged from the reactor to a gypsum slurry tank. The slurry is then pumped to the gypsum stack where gypsum settles from the slurry by gravity and the supernatant liquid is returned to the process.

ENGINEERING CHARACTERISTICS OF CT-121 FGD GYPSUM

Detailed laboratory and field testing of CT-121 FGD gypsum was performed as part of the overall stacking evaluation. Research emphasis was on assessing the physical and chemical properties, sedimentation-consolidation behavior, permeability characteristics, and shear strength characteristics of CT-121 FGD gypsum relevant to stacking methods of waste disposal. The effect of fly ash addition on the engineering behavior of CT-121 FGD gypsum was also briefly investigated.

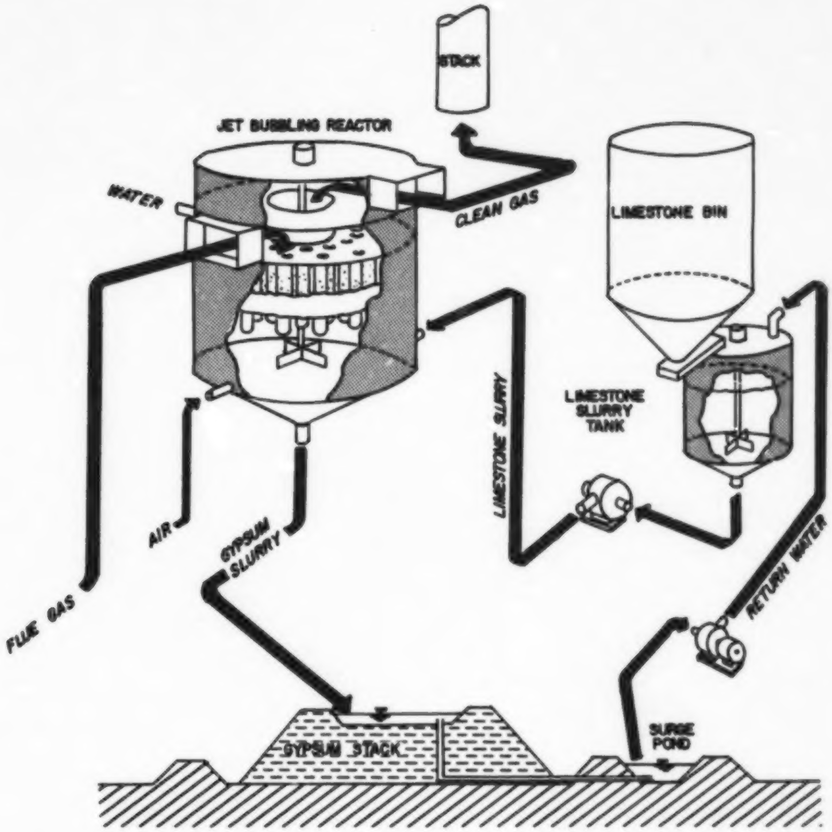

STACK
JET BUBBLING REACTOR
WATER
CLEAN GAS
LIMESTONE BIN
LIMESTONE SLURRY TANK
LIMESTONE SLURRY
AIR
GYPSUM SLURRY
FLUE GAS
RETURN WATER
SURGE POND
GYPSUM STACK

Figure 2. CT-121 Process Flow Diagram.

Mineralogical Analysis and Grain Size Distribution

X-ray diffraction data from random samples of the Japanese pilot plant and Plant Scholz CT-121 FGD gypsum both indicated that gypsum ($CaSO_4 \cdot 2H_2O$) was the only crystalline phase present. The X-ray diffraction traces were virtually identical to a trace obtained from analytical reagent grade gypsum.

The crystal structure and form of pilot plant and Plant Scholz CT-121 FGD gypsum are illustrated by scanning electron photomicrographs in Figure 3. As shown, the gypsum crystals are generally elongated with sharp, regular edges. The crystals vary in length from 0.05 to 0.25 mm with an average of about 0.13 mm, and vary in width from 0.03 to 0.06 mm.

Results from sieve and hydrometer analyses or the CT-121 FGD gypsum are presented in Figure 4. As shown, CT-121 FGD gypsum would be classified as a nonplastic, poorly-graded coarse silt.

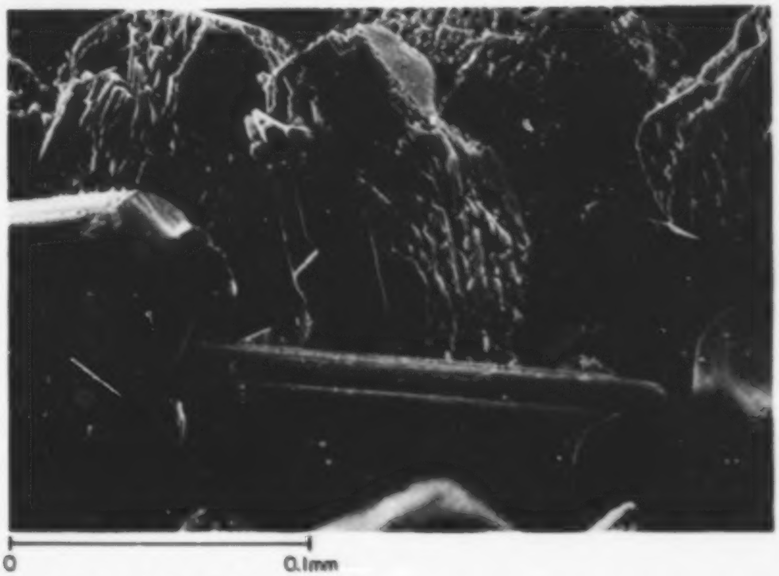
Sedimentation and Consolidation

The typical sedimentation-consolidation behavior of CT-121 gypsum is summarized in Figure 5. As shown, the initial void ratio and dry density after gravity sedimentation in gypsum-saturated water were 0.88 to 0.91 and 78.0 to 75.0 lb/ft^3 (71 to 73 percent solids), respectively.

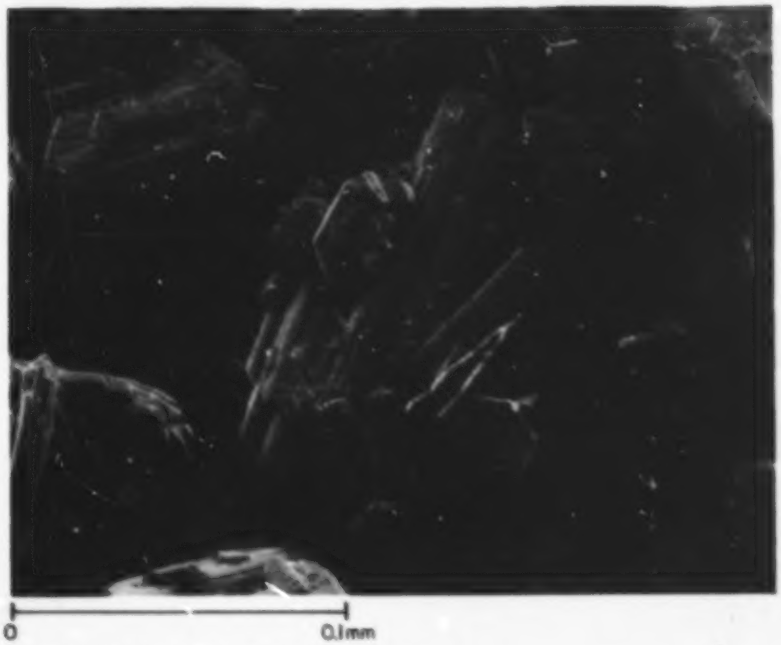
Pore fluid pH was found to have some effect on the initial void ratio and dry density. The initial void ratio of CT-121 FGD gypsum sedimented in untreated pH 3.0 gypsum-saturated water was consistently lower (about 5 to 7 percent) than the initial void ratio in neutralized pH 6.5 gypsum-saturated water. Therefore, the pore fluid pH will have some effect on the initial sedimented void ratio, but the affect will be relatively minor.

Consolidation of CT-121 FGD gypsum occurs quickly. Coefficients of consolidation typically range from 10 cm^2/sec at a stress of 0.01 kg/cm^2 (0.98 kPa) to greater than 200 cm^2/sec at stresses of 1.0 kg/cm^2 (98.1 kPa) and higher. These data indicate that CT-121 FGD gypsum will consolidate almost simultaneously with deposition for any reasonable rate of stack construction.

Based upon the study of the sedimentation and consolidation behavior of CT-121 FGD gypsum the following conclusions can be summarized relevant to gypsum stacking:



A. Plant Scholz FGD Gypsum



B. Chiyoda Pilot Plant Gypsum

Figure 3. Scanning Electron Photomicrographs of CT-121 FGD Gypsum.

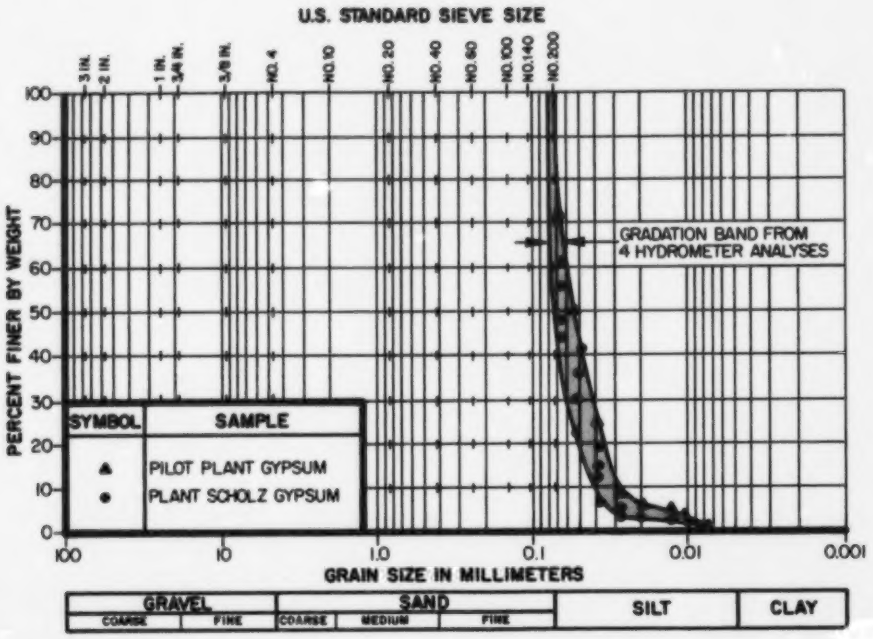


Figure 4. Grain Size Distribution of CT-121 FGD Gypsum.

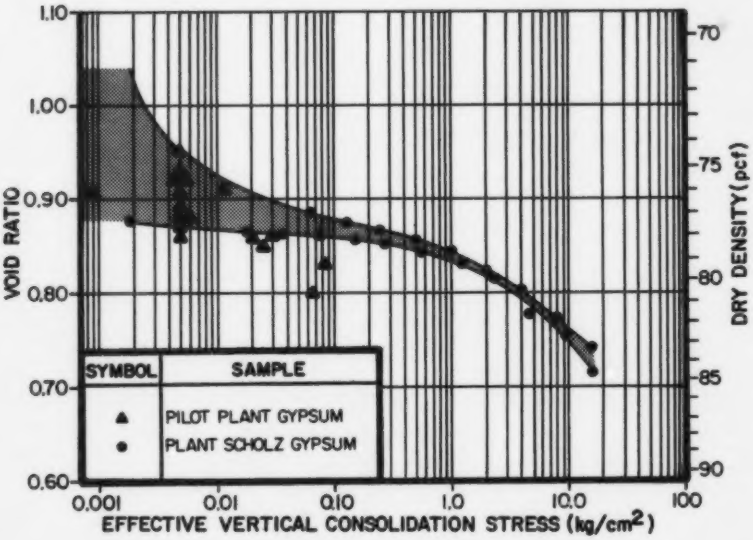


Figure 5. Sedimentation-Consolidation Behavior.

- CT-121 FDG gypsum settles rapidly. For the average crystal size of 0.06 mm equivalent diameter (see Figure 4), the settling velocity of 20 cm/min.

- After initial sedimentation or settling in gypsum-saturated liquor, to a dry density of 75 to 77 lb/ft^3 (71 to 72 percent solids), CT-121 FGD gypsum consolidates almost simultaneously with deposition. Depending on the height of the stack, the dry density of sedimented gypsum within the stack may increase an additional 5 to 15 lb/ft^3 (0.79 to 2.4 kN/m^3) to 80 to 90 lb/ft^3 (12.6 to 14.1 kN/m^3).

- The pH of the gypsum-saturated liquor has little effect on the sedimentation and consolidation behavior of CT-121 FGD gypsum.

Permeability

The effect of dry density and void ratio on the coefficient of permeability of sedimented CT-121 FGD gypsum is clearly illustrated in Figure 6. Laboratory sedimented gypsum samples were prepared to simulate sedimented gypsum within the stack. At dry densities less than 80 lb/ft^3 (12.6 kN/m^3), the coefficient of permeability for intact sedimented gypsum ranges from 1.0×10^{-3} cm/sec. As the dry density increases to 85 lb/ft^3 (13.3 kN/m^3), the coefficient of permeability for sedimented gypsum decreases to 6.0×10^{-4} cm/sec. Pore fluid pH was found to have no measurable effect on the coefficient of permeability for CT-121 FGD gypsum.

Cast gypsum samples were also prepared to simulate the casting of gypsum by a dragline during construction of the stack perimeter dike. The cast gypsum samples generally have a lower void ratio than the sedimented gypsum samples, and corresponding lower coefficients of permeability. The dry density of cast gypsum from the laboratory tests varies in the range of 85 to 95 lb/ft^3 (13.3 to 14.9 kN/m^3) with coefficients of permeability in the range of 2.0×10^{-4} to 5.0×10^{-4} cm/sec.

Shear Strength

Three objectives of the laboratory investigation of the shear strength characteristics of CT-121 FGD gypsum relevant to stacking were to determine:

- The relationship between dry density (or void ratio) and the effective friction angle.

- The effect of the pore fluid pH of gypsum-saturated liquor on the shear strength.

- The nature and magnitude of cohesion, if any, developing from cementation.

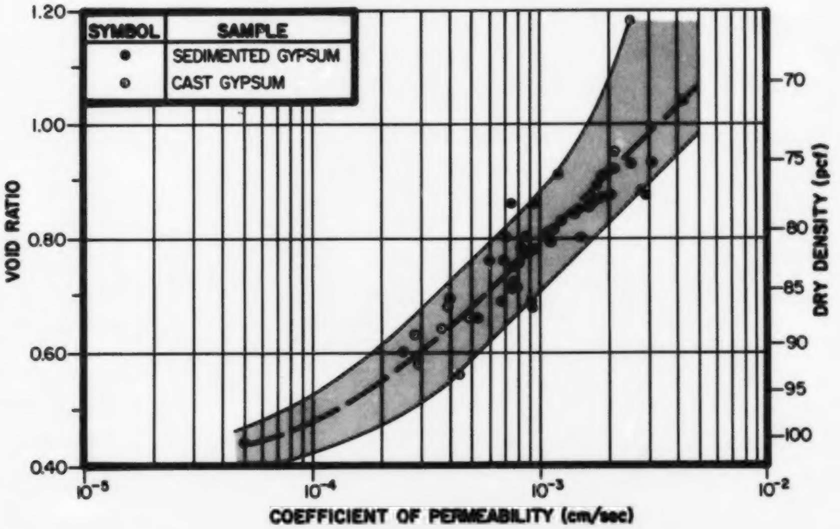


Figure 6. Void Ratio versus Coefficient of Permeability.

Three types of samples were tested: (1) laboratory sedimented gypsum to simulate the sedimentation of gypsum within the gypsum stack, (2) laboratory cast gypsum to simulate the casting of gypsum by a dragline during construction of the stack perimeter dike, and (3) undisturbed samples of cast gypsum from the Plant Scholz gypsum stack perimeter dike. The effect of pore fluid on shear strength was also investigated with pH 3.0 and 6.5 gypsum-saturated water, corresponding to untreated and neutralized gypsum slurry, respectively.

Figure 7 summarizes the effective friction angle versus void ratio and dry density for numerous cast and sedimented CT-121 FGD gypsum samples (from consolidated undrained triaxial compression tests with pore pressure measurements). These data indicate a friction angle increasing from 40.5° at a dry density of 78 lb/ft^3 (12.1 kN/m^3) to 46.5° at a dry density of 103 lb/ft^3 (16.2 kN/m^3). The range in dry density from 78 to 103 lb/ft^3 (12.2 to 16.2 kN/m^3) covers most densities likely to occur in a gypsum stack except for loose, initially sedimented gypsum at very low consolidation stresses.

Pore fluid pH was found to have no measurable effect on the shear strength of cast or sedimented gypsum. At similar void ratios, samples of CT-121 FGD gypsum with pH 3.0 and pH 6.5 pore fluid displayed essentially identical stress-strain-strength behavior.

An important strength characteristic typical of some phosphate gypsums relevant to stacking is the development of cohesion from cementation. The cementation is believed to develop because of the solubility of gypsum in acid or rain water, allowing dissolution and subsequent recrystallization as a result of seepage and evapotranspiration. Chemical constituents of the process acid waters are also believed to be a significant factor affecting cementation. These field conditions causing cementation, however, generally cannot be simulated into the laboratory. Laboratory samples of gypsum were allowed to dry under varying conditions in an attempt to simulate the drying which occurs in the perimeter dike. No true cohesion from cementation, however, was observed for any of these specimens upon resaturation and subsequent shearing in triaxial compression tests.

In summary, the following conclusions can be advanced concerning the shear strength of CT-121 FGD gypsum relevant to stacking:

- o The shear strength characteristics of CT-121 FGD gypsum are acceptable for stacking methods of waste disposal. The

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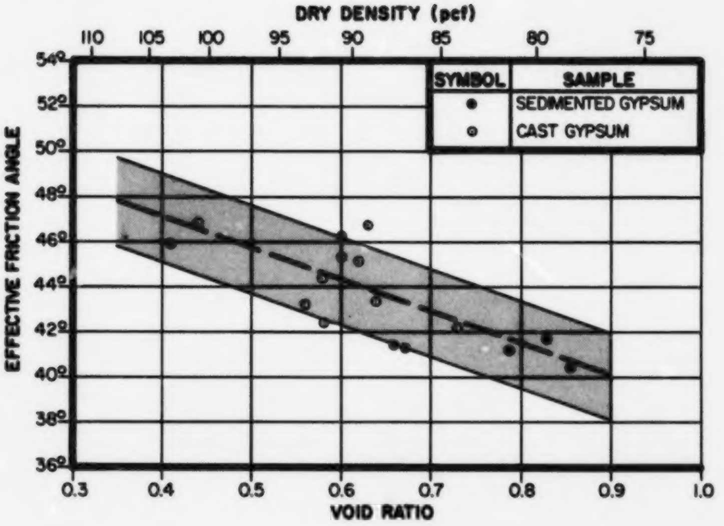

Figure 7. Void Ratio versus Effective Friction Angle.

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effective friction angle was generally found to increase from 40.5° at a dry density of 78 lb/ft^3 (12.2 kN/m^3) to 46.5° at a dry density of 103 lb/ft^3 (16.2 kN/m^3).

- The pH of the gypsum-saturated liquor has little effect on the shear strength of CT-121 FGD gypsum.
- No cohesion from cementation has been found to develop for CT-121 FGD gypsum for the laboratory and field conditions investigated during this study.

Effect of Fly Ash Addition on Engineering Characteristics

During one phase of the Chiyoda test program at Plant Scholz, fly ash was collected at the same time as CT-121 FGD gypsum. This occurred during particulate tests which allowed venturi liquor containing the fly ash to enter the Jet-Bubbling Reactor. The resulting slurry was piped to the stacking area as normally accomplished with gypsum alone. Since the potential exists for FGD systems to simultaneously require the disposal of fly ash and gypsum, the effect of fly ash addition on the stacking behavior of CT-121 FGD gypsum is of interest.

The results of laboratory testing on the effect of fly ash addition on the permeability, sedimentation-consolidation, and shear strength characteristics of CT-121 FGD gypsum all indicated reductions in the favorable stacking characteristics of CT-121 FGD gypsum. Since it was not within the scope of the CT-121 evaluation to establish the stackability of gypsum-fly ash mixtures, no definitive conclusions can be advanced on the potential for successfully stacking gypsum-fly ash mixtures. The laboratory tests, however, indicated the following conclusions:

- The strength characteristics of gypsum-fly ash mixtures appear satisfactory for stacking methods of waste disposal provided the mixtures are drained and sedimented.
- The lower dry density, higher water content, and lower coefficient of permeability of gypsum-fly ash mixtures in comparison to pure gypsum will definitely make the excavation of sedimented material and the casting of the perimeter dike more difficult.

Since the addition of fly ash to gypsum produces a material which is not as well suited for stacking as pure gypsum, the stacking of gypsum-fly ash mixtures should be avoided to obtain the greatest benefit and ease of construction from the favorable stacking characteristics of FGD gypsum. Research is necessary, however, to determine if feasibility alternatives exist for simultaneous disposal of fly ash and FGD gypsum.

PLANT SCHOLZ CT-121 FGD GYPSUM STACK

The objective of constructing and operating a prototype FGD gypsum stack was to establish the feasibility of disposing of CT-121 FGD gypsum by stacking with a dragline using the upstream method of construction. Although by-product gypsum from the manufacture of phosphate fertilizer has been successfully stacked, no field experience was available on the stacking characteristics of gypsum produced by FGD scrubbers.

Stacking waste gypsum normally utilizes the upstream method of construction. In this method, illustrated in Figure 8, an earthen starter dike is first constructed to form a sedimentation pond and stacking area. Gypsum is then pumped to the sedimentation pond in slurry form, usually at 10 to 20 percent solids, and allowed to settle and drain. Process water is decanted from the pond and returned to the plant. Once sufficient gypsum is deposited within the pond, gypsum is excavated with a dragline to raise the perimeter dikes of the stack. The draglines typically have a working reach of 60 feet and a 2 to 3 cubic yard bucket. The process of sedimentation, excavation, and raising of the perimeter dikes continues on a regular basis during the active life of the stack. An illustration of a typical stack design is shown in Figure 9. It is noteworthy to identify that the sedimentation pond is typically divided into two parts, such that one area can be receiving slurry while the other compartment is drained prior to excavation.

Stack Construction and Operation

The structural and environmental performance of the stack was monitored over a nine-month test period from October 1978 through June 1979, with additional field investigations and evaluations continuing through June 1980. Although the prototype stack is small (i.e., one-half acre (2023 m^2) and 12 feet (3.7 m) high, the study allowed a comparison of laboratory tests of geotechnical engineering properties relevant to stacking with actual field behavior and a comparison of the stacking performance of CT-121 FGD gypsum with gypsum produced by the phosphate fertilizer industry.

The disposal area and gypsum stack were proportioned for an estimated nine-month gypsum production of 5,500to 6,500 tons (5.0×10^6 to 5.9×10^6 kg) and a final stack height of 25 feet (7.6 m). Actual gypsum production during the test program, however, required a reduction in the final stack height. The stack geometry was also governed by the minimum dimensions required for: (1) safe operation of the

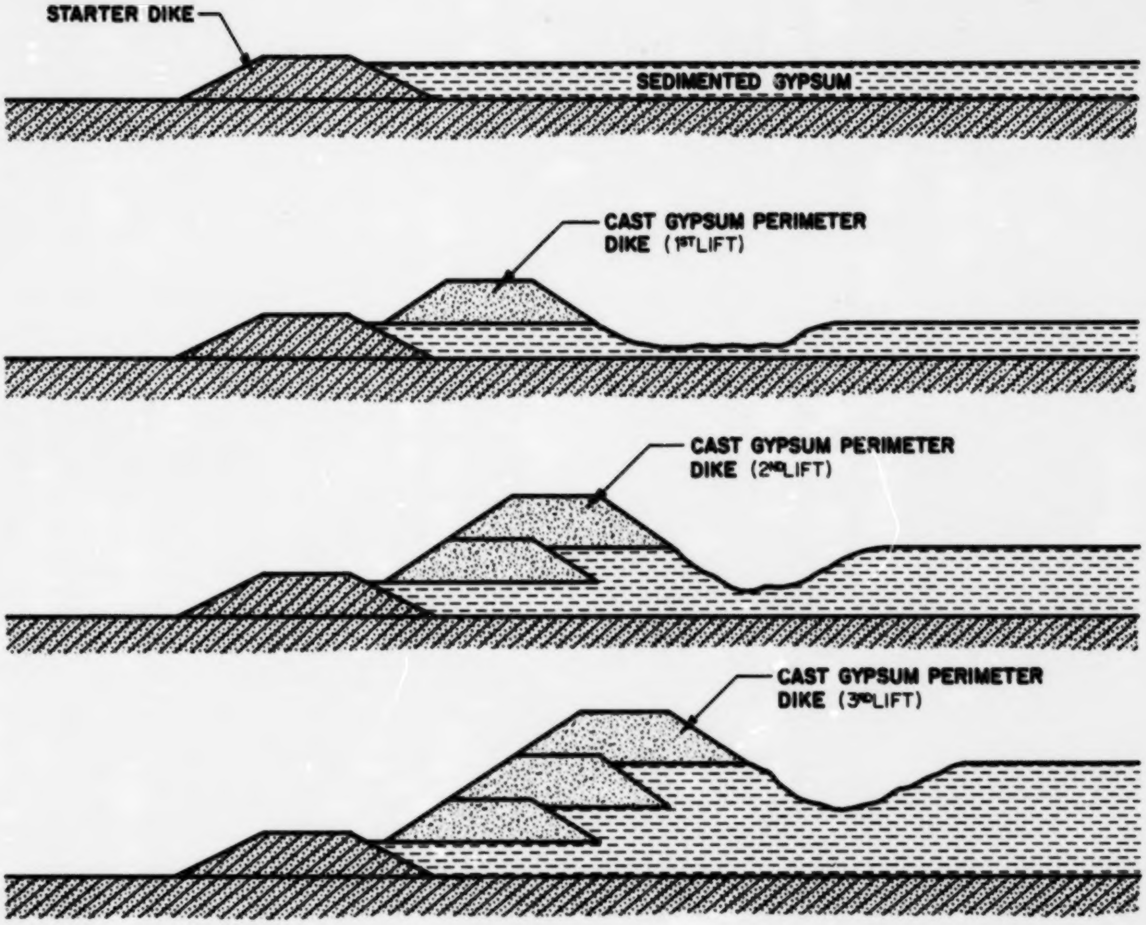
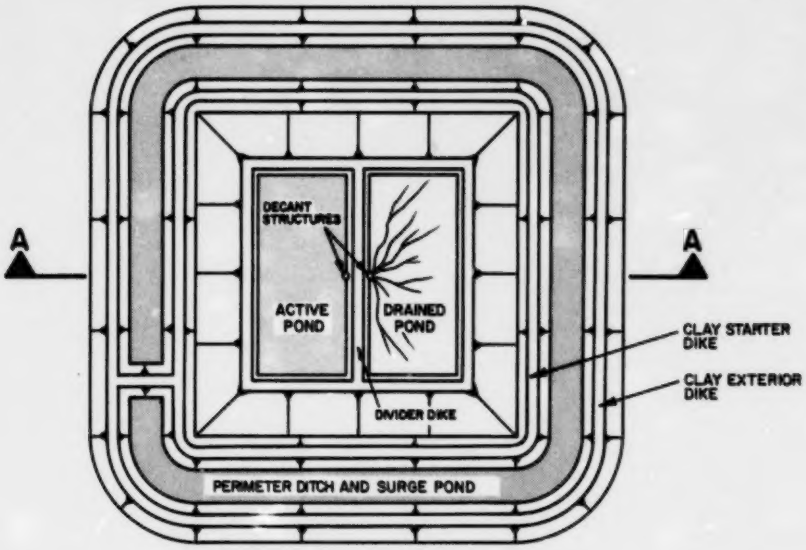

STARTER DIKE
SEDIMENTED GYPSUM
CAST GYPSUM PERIMETER DIKE (1ST LIFT)
CAST GYPSUM PERIMETER DIKE (2ND LIFT)
CAST GYPSUM PERIMETER DIKE (3RD LIFT)

Figure 8. Upstream Method of Gypsum Stack Construction.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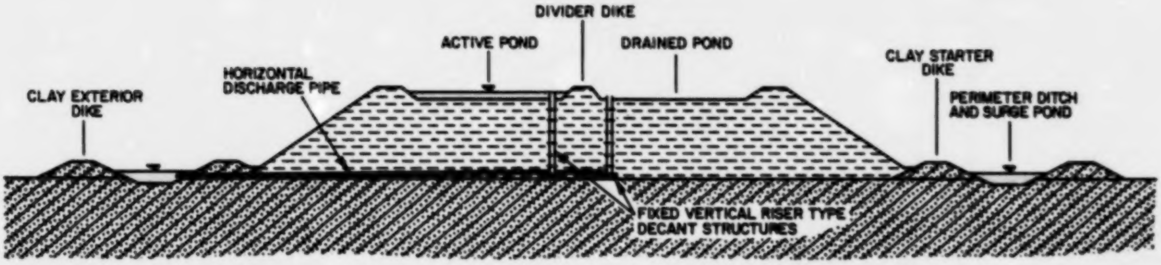
A
DECANT STRUCTURES
ACTIVE POND
DRAINED POND
CLAY STARTER DIKE
CLAY EXTERIOR DIKE
DIVIDER DIKE
PERIMETER DITCH AND SURGE POND

GYPSUM STACK SITE PLAN


DIVIDER DIKE
ACTIVE POND
DRAINED POND
CLAY STARTER DIKE
CLAY EXTERIOR DIKE
HORIZONTAL DISCHARGE PIPE
PERIMETER DITCH AND SURGE POND
FIXED VERTICAL RISER TYPE DECANT STRUCTURES

SECTION A-A GYPSUM STACK CROSS SECTION

Figure 9. Typical Gypsum Stack Design.

dragline from the perimeter dike of the stack and (2) provision of sufficient storage capacity within the center of the stack to gypsum and allow clarification of the process water. For the small quantity of gypsum scheduled to be produced during the nine-month test program, the stack also had to be small in area such that the stack could be raised to a height sufficient to allow observation the stacking behavior of CT-121 FGD gypsum and slope stability of the cast gypsum dikes.

The selected site plan and typical cross section of the disposal area and gypsum stack are shown in Figure 10. The stacking area was located adjacent to the north side of an existing settling pond. The disposal area encloses approximately 0.5 acres (2023 m^2) within 375 feet (114 m) of starter dike. Figure 11 illustrates the location of the gypsum stack at Plant Scholz.

A liner was not installed within the gypsum stacking area because the underlying soils were thought to be sufficiently impervious to prevent any significant migration of leachate from the stacking area. In addition, the stack was only to be used for a short period of time before retirement and eventual removal.

Gypsum and process water were initially deposited in the stacking area on October 12, 1978. Process water was pumped to the stacking area at rates varying from 30 to 70 gal/min (0.0019 to 0.0028 m^3/sec) and solids contents varying from 5 to 15 percent. Gypsum production varied considerably during the nine-month test program within the range of 12 to 26 tons per day (10,900 to 23,600 kg/day).

The gypsum stack was raised a total of four times by the upstream method of construction (as shown in Figure 8) during the nine-month test program. Figure 12 presents photographs of the disposal area prior to the initial raising of the gypsum stack. As shown, the sedimented gypsum can be walked upon at most locations within the containment area except near the slurry outlet where the sedimenting gypsum is still very loose.

Figure 13 presents photographs taken during construction of portions of the cast dike and perimeter ditch on the north side of the stack. As shown, the dragline first excavates gypsum from the containment area adjacent to the starter dike to form the perimeter ditch. The gypsum excavated from the perimeter ditch is then cast to form the perimeter dike of the gypsum stack. Photograph 13A shows the dragline removing gypsum from the perimeter ditch and casting the material to form the stack perimeter dike. Photograph 13B shows the perimeter ditch and cast dike

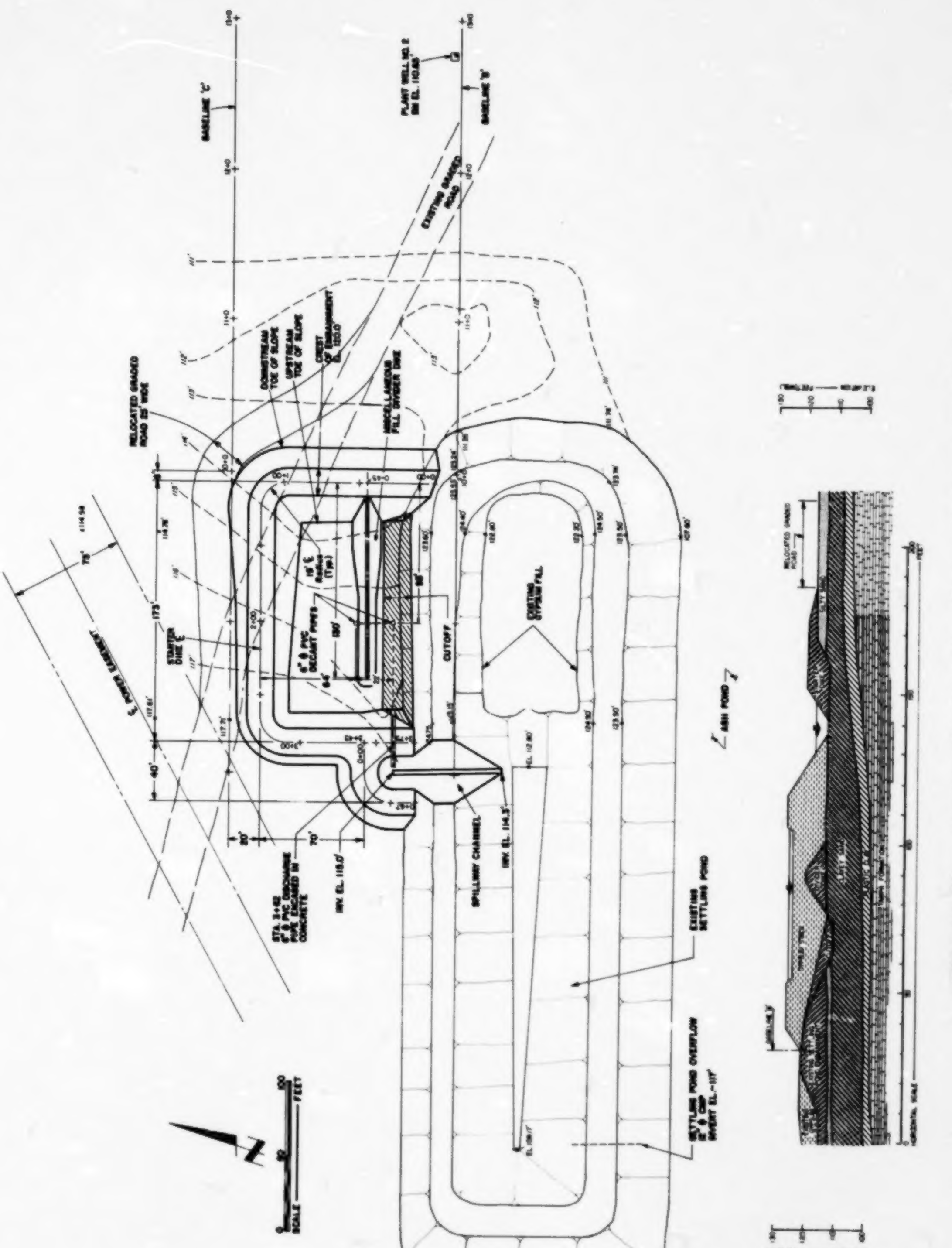

RELOCATED GRADED ROAD 25' WIDE
DOWNSTREAM TOE OF SLOPE
UPSTREAM TOE OF SLOPE
CREST OF EMBANKMENT EL. 120.0'
MISCELLANEOUS FILL DIVIDER DIKE
EXISTING GRADED ROAD
BASELINE 'C'
BASELINE 'B'
PLANT WELL NO. 2
STARTER DIKE
6" Ø PVC DECANT PIPES
OUTFALL
EXISTING GYPSUM FILL
ASH POND
SPILLWAY CHANNEL
INV. EL. 114.5'
INV. EL. 118.0'
EXISTING SETTLING POND
SETTLING POND OVERFLOW
SCALE
FEET
RELOCATED GRADED ROAD
BASELINE 'B'
HORIZONTAL SCALE
ELEVATION

Figure 10. Gypsum Stack Site Plan and Cross-Section.

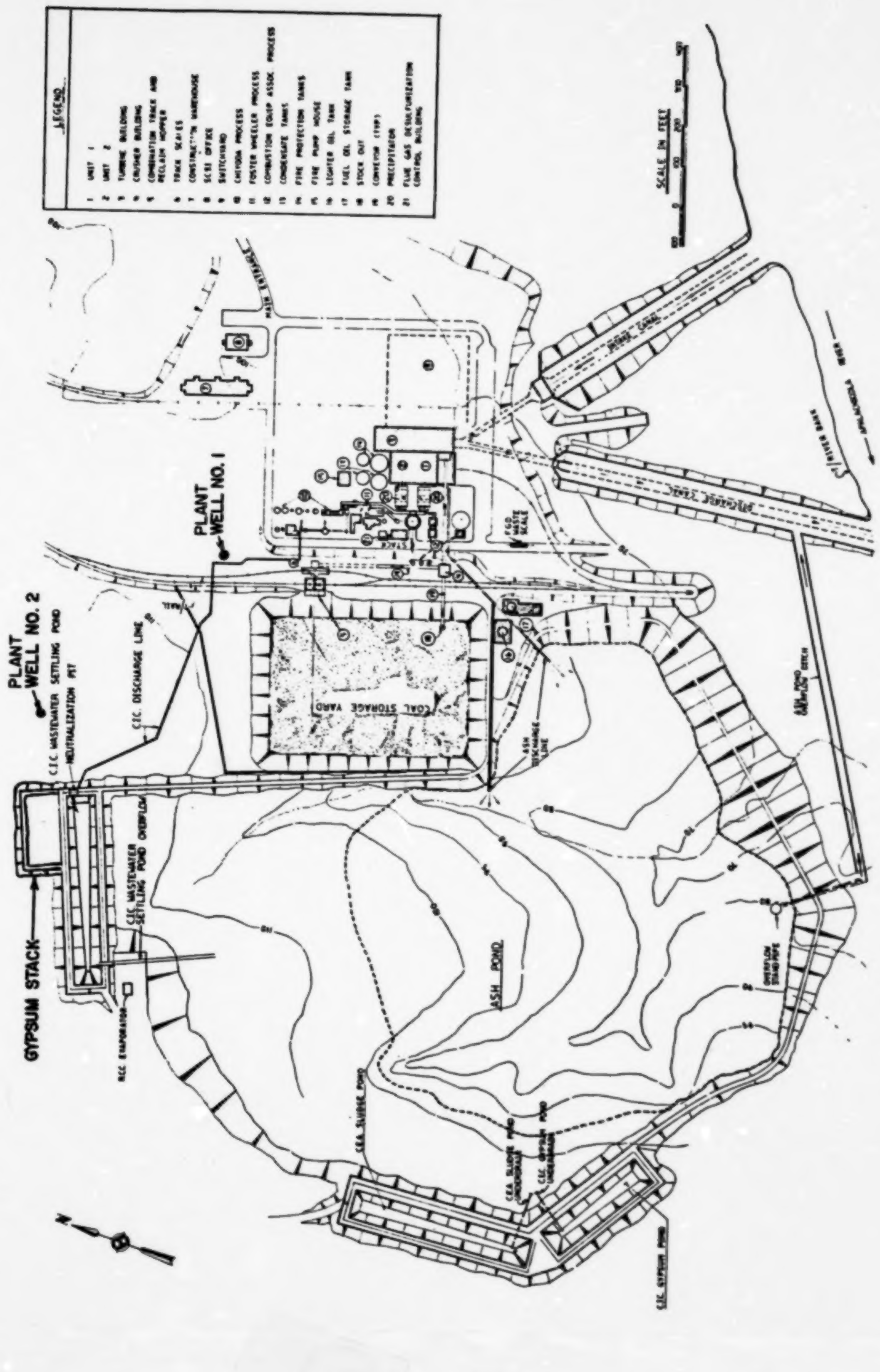

LEGEND
1 UNIT 1
2 UNIT 2
3 TURBINE BUILDING
4 CRUSHER BUILDING
5 COMBINATION TRACK AND RECLAIM HOPPER
6 TRUCK SCALES
7 CONSTRUCTION WAREHOUSE
8 SCSI OFFICE
9 SWITCHYARD
10 CHIYODA PROCESS
11 FOSTER WHEELER PROCESS
12 COMBUSTION EQUIP ASSOC. PROCESS
13 CONDENSATE TANKS
14 FIRE PROTECTION TANKS
15 FIRE PUMP HOUSE
16 LIGHTER OIL TANK
17 FUEL OIL STORAGE TANK
18 STOCK OUT
19 CONVEYOR (TYP.)
20 PRECIPITATOR
21 FLUE GAS DESULFURIZATION CONTROL BUILDING
SCALE IN FEET
100 0 100 200 300 400 500
MAIN ENTRANCE
INTAKE CANAL
DISCHARGE CANAL
RIVER BANK
APALACHICOLA RIVER
PLANT WELL NO. 1
PLANT WELL NO. 2
STACK
F.G.D. WASTE SCALE
GYPSUM STACK
C.I.C. WASTEWATER SETTLING POND NEUTRALIZATION PIT
C.I.C. DISCHARGE LINE
C.I.C. WASTEWATER SETTLING POND OVERFLOW
R.C.C. EVAPORATOR
COAL STORAGE YARD
ASH DISCHARGE LINE
ASH POND
ASH POND RUNOFF DITCH
OVERFLOW STANDPIPE
C.E.A. SLUDGE POND
C.E.A. SLUDGE POND UNDERDRAIN
C.I.C. GYPSUM POND UNDERDRAIN
C.I.C. GYPSUM POND
N

Figure 11. Scholz Electric Generating Station Site Plan.

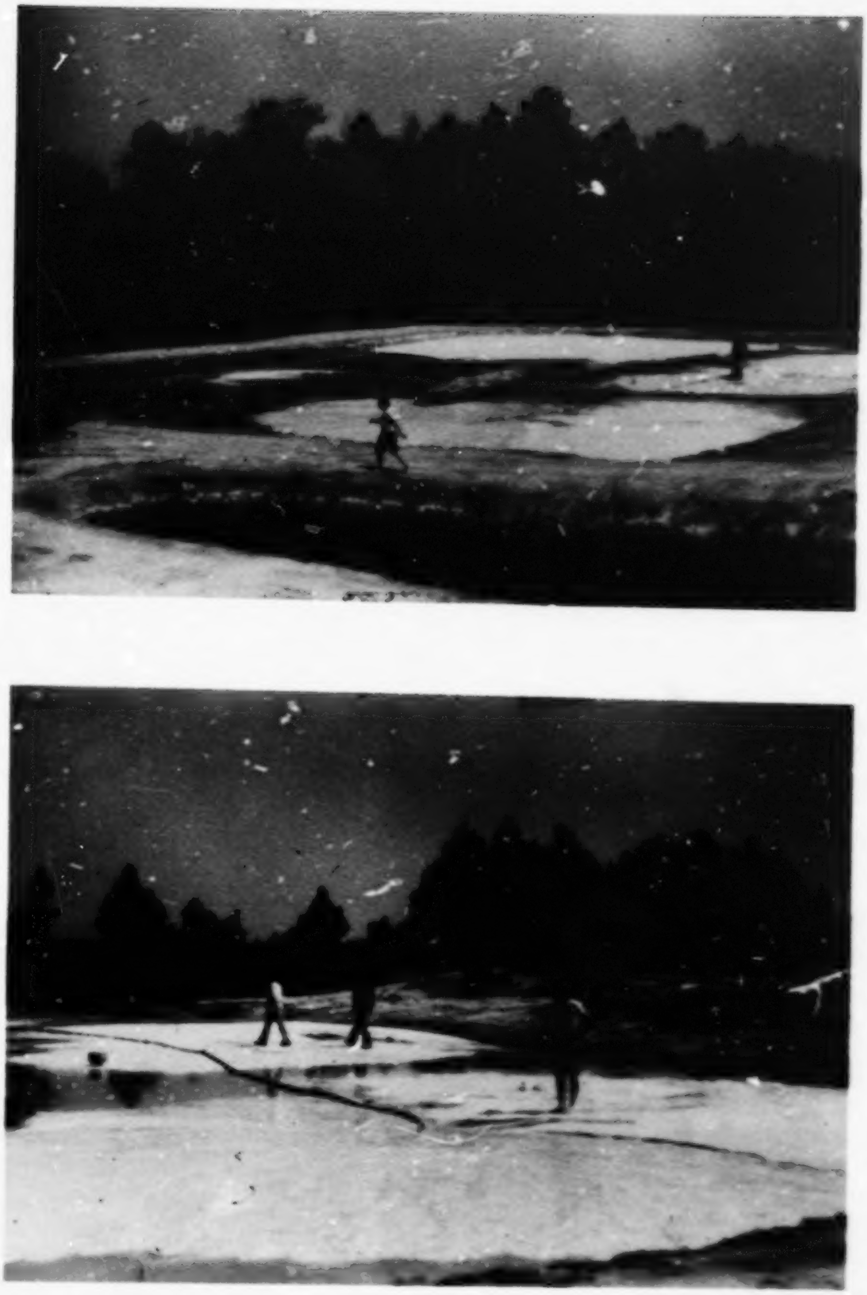


Figure 12. Stacking Area Prior to Initial Dike Raising.

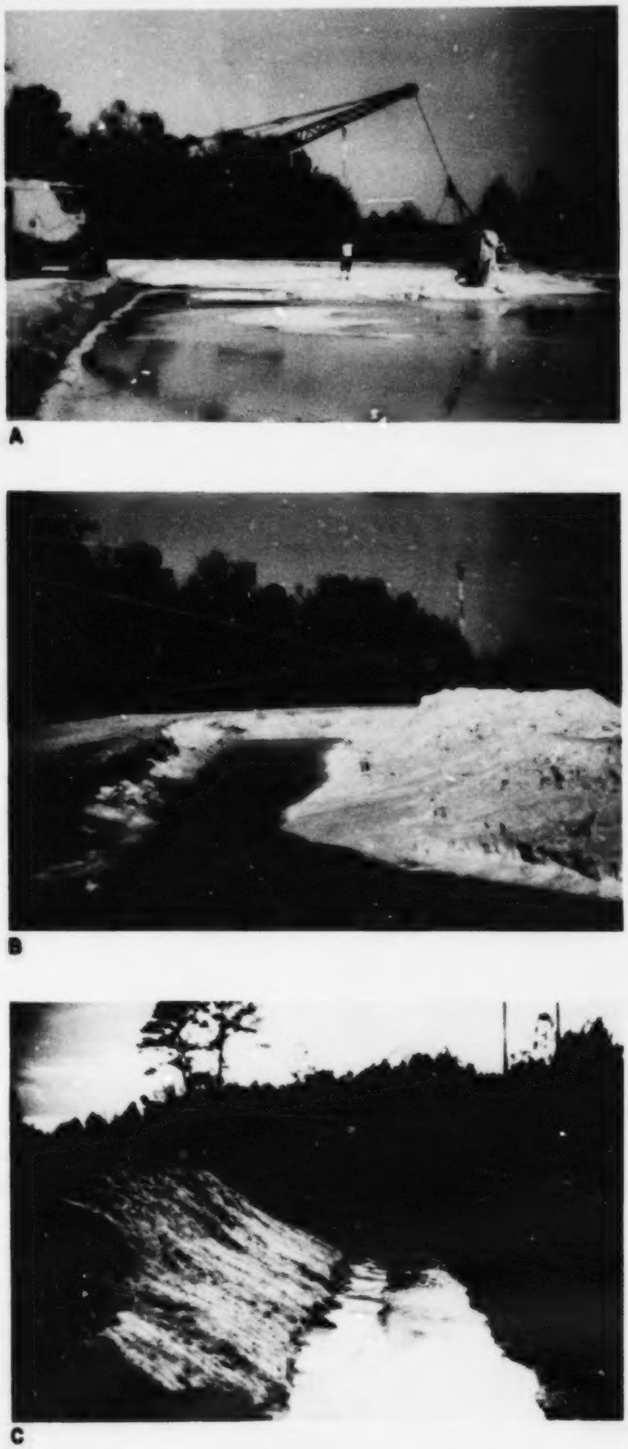


Figure 13. Raising North Wall of Cast Gypsum Dike.

shortly after construction. At this time, the gypsum was still too wet to cast more than 2 to 3 feet (0.6 to 0.9 m) high without allowing some time for draining and drying. After drainage and drying from exposure to the sun, the gypsum was cast much easier and at steeper slopes as shown in Photograph 13C. As shown in Figure 14, the cast gypsum dikes of the stack were sufficiently stable and trafficable to allow the dragline to work upon the dikes with no difficulty.

The gypsum slurry was generally discharged at the corners of the stack through a 3-inch diameter flexible hose and allowed a flow toward the decant pipe at the center of the stacking area. The slurry was discharged at the corners of the stack to provide the longest flow path, and hence the most efficient clarification of the process water. Since the stacking area was reduced significantly in size after raising the dikes, complete clarification of the process water became difficult and some gypsum was carried over through the decant pipe. The loss of gypsum, however, was not a major problem. The decant pipe was raised as necessary in approximately 1-foot increments as gypsum sedimented within the stack.

The sedimentation of gypsum within the stack progressed satisfactorily, except for an initial problem with some sloughing along the base of the downstream slope of the perimeter dike. The seepage induced sloughing resulted from the absence of significant cohesion within the steep outer scope of the gypsum perimeter dike and was aggravated by the relatively thin width of the perimeter dike in comparison to the head difference across the dike. The sloughing generally extended 2 to 3 feet above the water level in the perimeter ditch. When the perimeter dike was raised for the second time, the downstream slopes were flattened near the base of the stack and the width of the perimeter dike was increased by several feet. No major sloughing occurred after the dike was raised for the second time.

Photographs of the completed gypsum stack are shown in Figure 15. The stack is approximately 100 feet (31 m) square at the top. The average height of the stack above the invert of the perimeter ditch is approximately 12 feet (3.7 m). The perimeter dike crest width varies from 9 to 15 feet (2.7 to 4.6 m) and the exterior slopes of the stack are generally 1.0 Vertical to 1.5 Horizontal. Photographs 15A and 15B show, respectively, the entire stack and north wall approximately one month after the process shutdown. At this time, process water from the settling pond was being pumped to the stack to maintain seepage through the stack. Seepage was maintained through the stack for a period of approximately one month after shutdown to further observe the effects of steady-state seepage on slope stability. Photograph 15C, taken 5 months after the process shutdown, shows

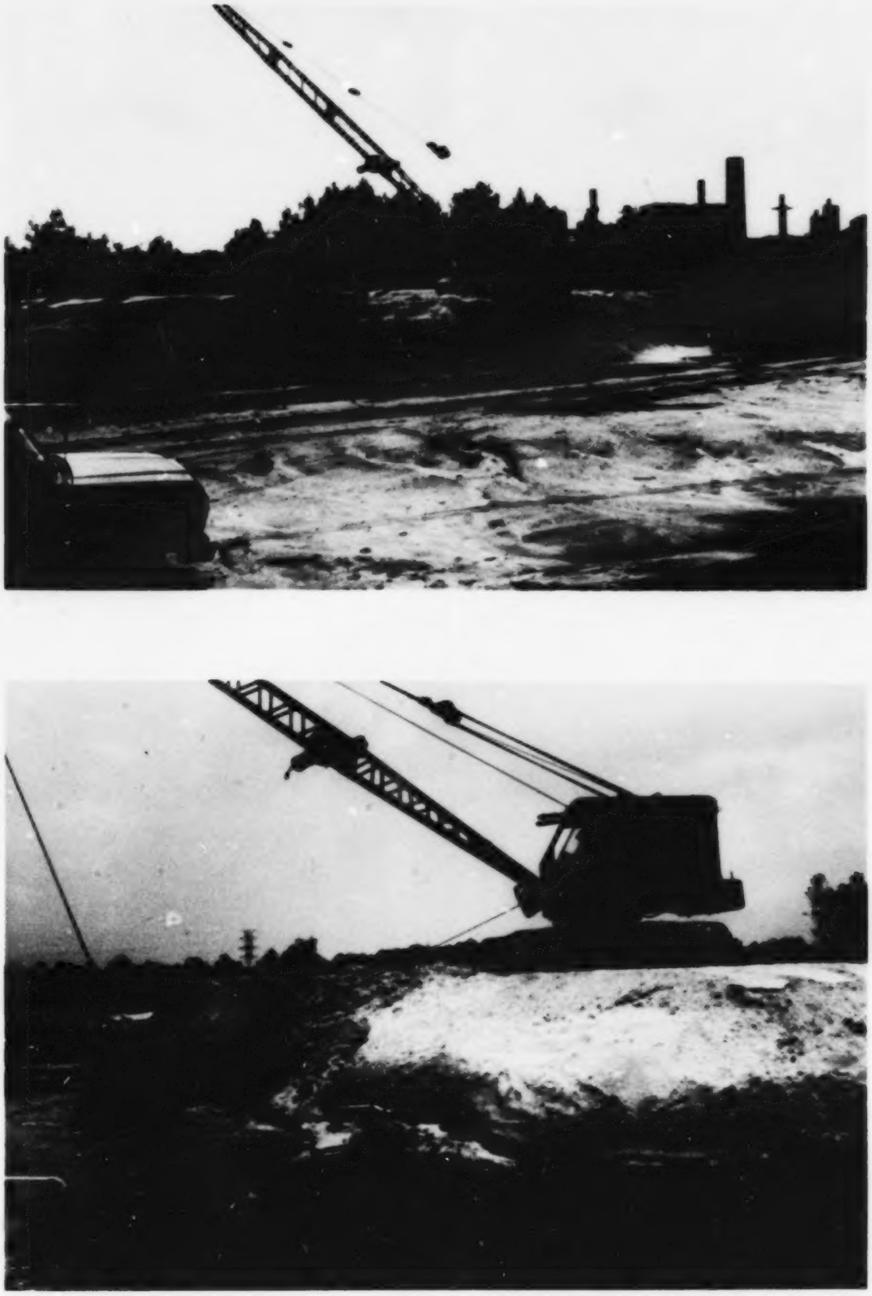


Figure 14. Dragline Operation From Crest of Cast Gypsum Dike.

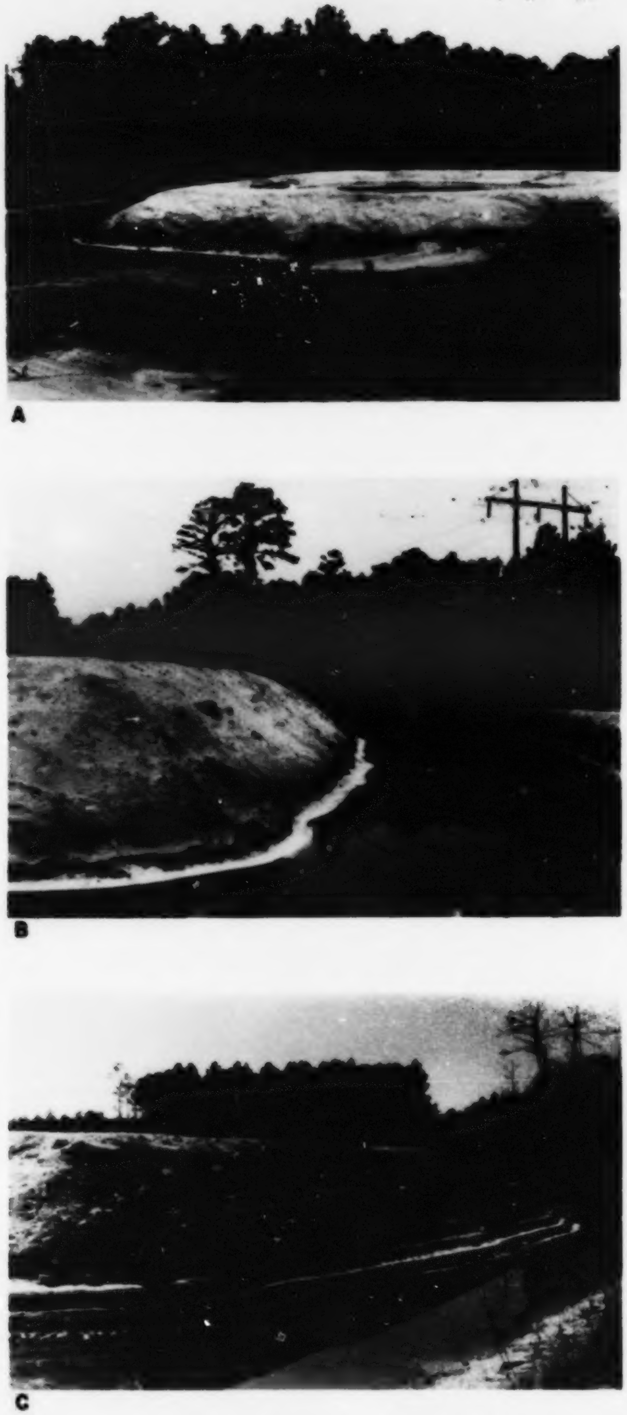


Figure 15. Completed Gypsum Stack.

no significant change in the appearance of the north wall after several months of weathering and aging.

Following completion of physical testing to determine aging and weathering effects, the gypsum stack at Scholz will be removed and sold. Gypsum stacks in the phosphate industry that have been retired have generally not been reclaimed. Probable future regulations (i.e., Resources Conservation and Recovery Act (RCRA) of 1976), however, may require regrading of the stacks to enhance runoff and minimize leachate, placement of a relatively impervious soil cover above the gypsum to minimize leachate, and reclamation with an indigenous vegetation (4,5).

Stacking Performance

Stacking of saturated CT-121 FGD gypsum from beneath the water surface of the undrained pond was possible, provided the gypsum was cast upon a dry surface and sufficient time was allowed for water to drain from the cast material before attempting to pile the gypsum more than 2 to 3 feet (0.6 to 0.9 m) high. Dry gypsum located above the water surface, which would be similar to gypsum within a drained pond, was excavated and cast easily and displayed acceptable stacking behavior.

Since no cohesion from cementation developed during the test program, sloughing often occurred below the springline (point where water emerges from outer surface of stack) for the relatively steep slopes of the cast gypsum dike. The overall stability of the slope, however, was generally not affected by the sloughing. In full-scale FGD gypsum stacks, flat slopes below the springline or internal drains may be required to control seepage induced instability and sloughing if some cohesion from cementation does not develop.

A fly ash-gypsum mixture deposited in the stack during particulate tests could be excavated with a dragline, although the process was much more time consuming. The fly ash-gypsum mixture could be cast although the poor drainage characteristics and high water content of the mixture generally produced much flatter cast slopes than occur with gypsum. Overall, the field performance indicated the addition of fly ash to CT-121 FGD gypsum significantly reduced the favorable stacking characteristics of CT-121 FGD gypsum. Laboratory tests on the effect of fly ash addition on the engineering characteristics of CT-121 FGD gypsum also indicated a general reduction in the favorable stacking characteristics of CT-121 FGD gypsum by the addition by fly ash.

As with phosphate gypsum stacks, the cast CT-121 FGD gypsum dikes and slopes developed a thin, hard drying crust. This crust apparently resulted from the dissolution of gypsum crystals from rainfall and subsequent recrystallization and drying. The slopes displayed essentially no erosion from rainfall. Dusting also was not a problem. Therefore, it is not expected that erosion protection will be required on the outside slope of FGD gypsum stacks. If long-term maintenance and reclamation require that the slopes be grassed, it may be expedient to flatten the slopes to 2.5 Horizontal to 1.0 Vertical or flatter as the stack is raised. Some small clumps of grass were observed growing on the stack slopes, and grassing with or without a topsoil dressing may be possible, although no research on this topic was performed with the CT-121 gypsum stack.

Groundwater Monitoring

Observation wells and piezometers were installed around the gypsum disposal area to monitor changes in groundwater quality during and after construction and operation of the gypsum stack. Since no "impermeable" liner was installed within the gypsum stacking area, some impact on the aquifers was anticipated. Due to the relatively impervious nature of the underlying soils and temporary nature of the stack, however, the impact was expected to be acceptable and locally isolated immediately below and adjacent to the stack. Accordingly, the observation wells and piezometers were installed close to the stack to detect changes in groundwater quality at the earliest possible time. The locations and formation penetrated by the observation wells and piezometers are shown in Figure 16. The locations of the wells and piezometers were selected to monitor the water quality both upstream and downstream of the stack relative to the direction of groundwater flow within both the surficial and Floridan aquifers.

Background water quality samples were collected from each observation well and piezometer prior to the placement of gypsum or process water within the stacking area. Subsequent water quality samples were obtained approximately once per month from each observation well and piezometer during the active life of the stack, and for one year beyond the active life of the stack, at six-month intervals.

The chemical analyses included major species, pH, total dissolved solids and conductivity. Trace elements within the groundwater, which may have entered the process water from the limestone and/or fly ash removed in the pre-scubber and Jet-Bubbling Reactor, were also monitored periodically as part of the overall

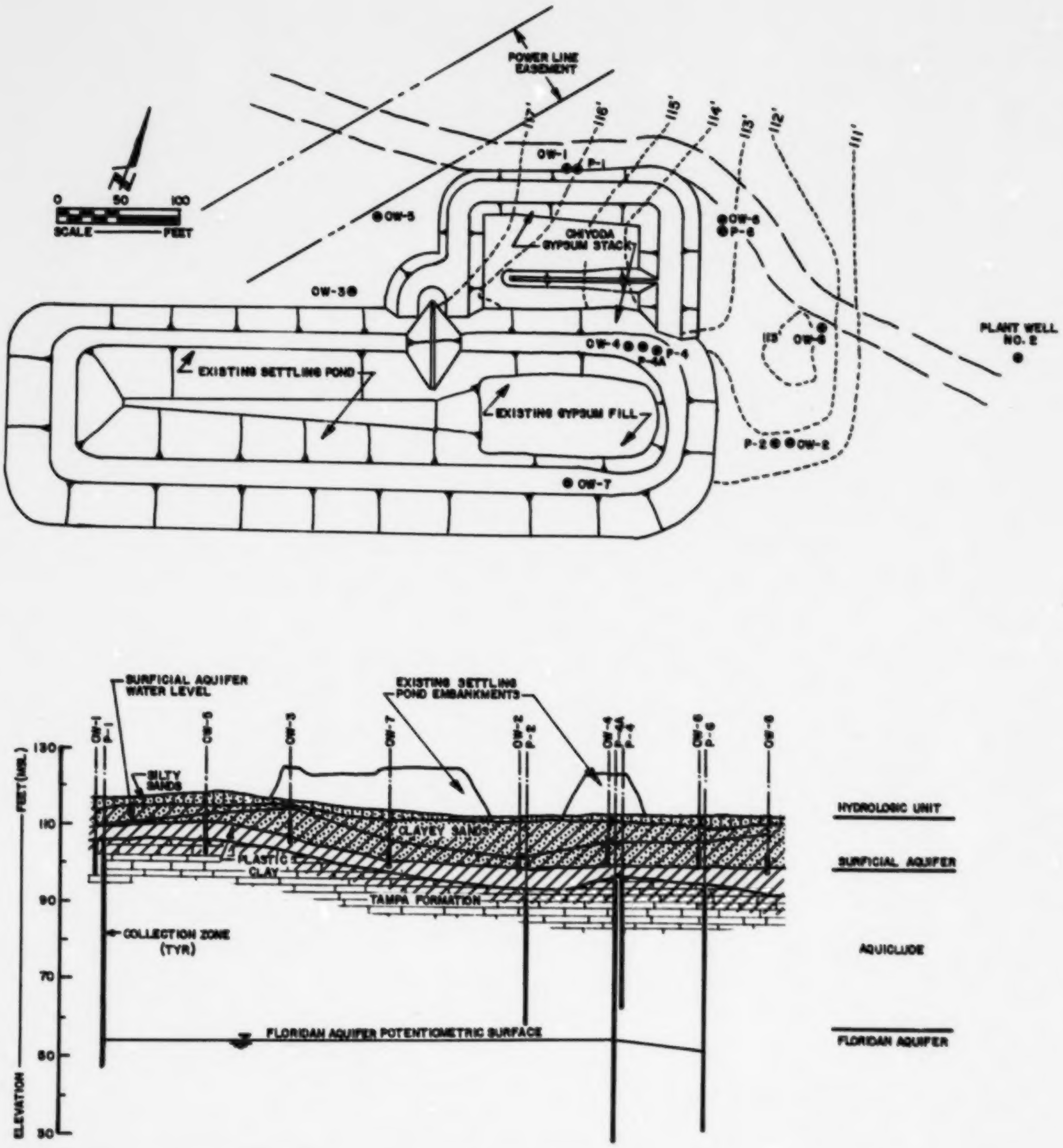

POWER LINE EASEMENT
117'
116'
115'
114'
113'
112'
111'
113'
N
0
50
100
SCALE FEET
OW-1
P-1
OW-5
OW-6
P-6
CHIYODA GYPSUM STACK
OW-3
OW-4
P-4
P-4A
OW-8
PLANT WELL NO. 2
EXISTING SETTLING POND
EXISTING GYPSUM FILL
P-2
OW-2
OW-7
SURFICIAL AQUIFER WATER LEVEL
EXISTING SETTLING POND EMBANKMENTS
OW-1
P-1
OW-5
OW-3
OW-7
OW-2
P-2
OW-4
P-4A
P-4
OW-6
P-6
OW-8
130
110
90
70
50
30
ELEVATION FEET (MSL)
SILTY SANDS
CLAYEY SANDS
PLASTIC CLAY
TAMPA FORMATION
COLLECTION ZONE (TYP)
FLORIDAN AQUIFER POTENTIOMETRIC SURFACE
HYDROLOGIC UNIT
SURFICIAL AQUIFER
AQUICLUDE
FLORIDAN AQUIFER
0
200
400
600
800
HORIZONTAL DISTANCE FEET

Figure 16. Groundwater Monitoring Location Plan and Profile.

groundwater monitoring program. Analyses required by the Florida Department of Environmental Regulation included monthly determinations of sulfate, calcium, sodium, pH and conductivity.

The process water reaching the disposal area was monitored on a monthly basis during the life of the gypsum stack. Results from these analyses are summarized in Table 1. The process water is a neutralized gypsum-saturated liquor and, therefore, contains high concentrations of calcium and sulfate ions. The process water is also high in chloride, magnesium, nitrate and sodium ions.

Table 1

PROCESS WATER CHEMICAL COMPOSITION

Parameter	Average Test Period Value
pH	7.4
Ca^{++}	740 mg/l
Mg^{++}	780 mg/l
Na^{+}	90 mg/l
Cl^{-}	890 mg/l
$SO_4^{=}$	3050 mg/l
NO_3^{-}	530 mg/l
TDS	890 mg/l

Several significant observations were apparent in water quality data obtained during the nine-month active life of the stack and five-month period following the process shut-down:

- No consistent increase in trace elements occurred in either the surficial or Floridan aquifer. Levels of arsenic, chromium and selenium, the trace elements of major pollution concern with the CT-121 FGD process liquor, were generally within acceptable drinking water standards.
- Leachate has not affected water quality without the deeper units of the Floridan aquifer as evidenced by no change from background conditions at Plant Well No. 2 within the Suwannee limestone (Figure 17A).
- Leachate has entered the Floridan aquifer within the upper unit of the Tampa formation immediately below the gypsum stack as evidence by consistent increases in all monitored parameters within Piezometers P-4S and P-6 (Figure 17B).

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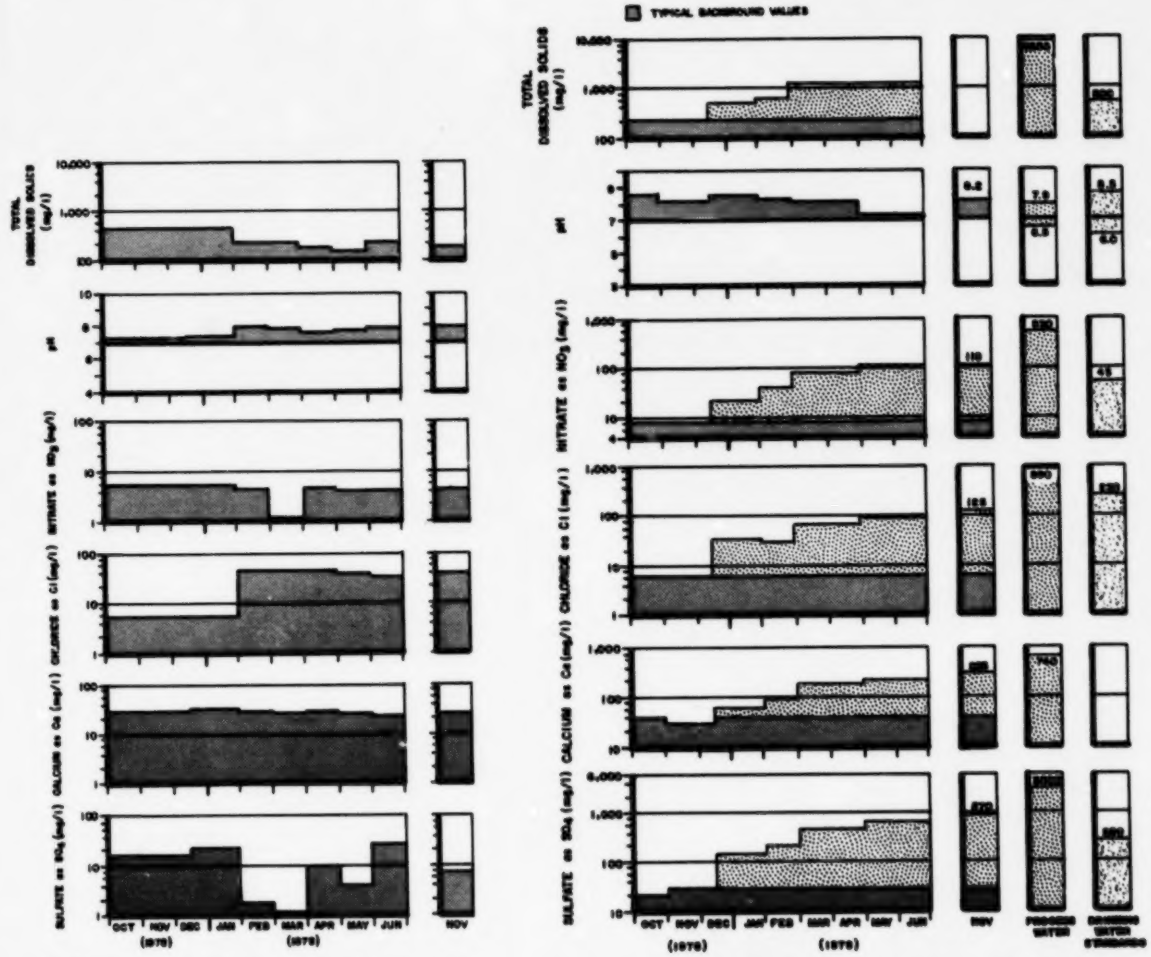

A. Floridan Aquifer Water Quality at Plant Well No. 2

B. Floridan Aquifer Water Quality at Piezometer P-4A and P-6

Figure 17. Floridan Aquifer Water Quality.

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- The surficial aquifer immediately adjacent to the gypsum stack in the direction of ground water flow (i.e., southeast of the gypsum stack) shows contamination in observation well OW-6 and OW-2, as evidenced by increases in all monitored parameters.

As with phosphate gypsum stacks, groundwater contamination from the CT-121 FGD scrubber process waters is a concern. The CT-121 process waters contain concentrations of sulfate, calcium, chloride, nitrate, magnesium and sodium that are several orders of magnitude greater than drinking water standards as well as natural background levels within the aquifers at Plant Scholz. Trace elements such as arsenic, chromium and selenium are also present within the process water at concentrations above drinking water standards and could pose a possible problem. For the stacking of CT-121 FGD gypsum to be environmentally acceptable over the long-term operation of a full-scale stack, the seepage of leachate should be controlled or prevented.

UTILIZATION TESTING

The potential utilization of by-product CT-121 FGD gypsum in the manufacture of wallboard and Portland cement, and as an agricultural soil amendment, were evaluated by Chiyoda International Corporation as part of the overall CT-121 process evaluation.

Two full-scale production demonstrations were conducted by U.S. Gypsum Company, Jacksonville, Florida and National Gypsum Company, Tampa, Florida utilizing approximately 100 tons of CT-121 FGD gypsum each. Their results indicated that CT-121 FGD gypsum could be used to manufacture wallboard to a quality equivalent to that obtained from natural virgin gypsum. Some minor material handling problems occurred, however, with the FGD gypsum, since the water content of the CT-121 FGD gypsum from the gypsum stack was slightly higher than that of virgin gypsum.

Gypsum is often used to retard the setting time of Portland cement. Laboratory tests performed by Flinkote Company, Calavaras Cement Division confirmed on a laboratory scale that CT-121 FGD gypsum is as acceptable as virgin gypsum for use as a retarder in Portland cement.

No agricultural utilization testing was planned for CT-121 FGD gypsum, although such tests were previously conducted by the University of Florida Agricultural Research and Education Center in Quincy, Florida on a limited basis with CT-101

FGD gypsum, a chemically and physically identical predecessor to CT-121 FGD gypsum (1). Findings from this previous study indicated that CT-101 FGD gypsum was a viable source of calcium and sulfur for peanut and soybean crops.

SUMMARY AND CONCLUSIONS

The feasibility of stacking CT-121 FGD gypsum has been evaluated using geotechnical laboratory testing to determine engineering properties relevant to stacking, and by observation of a prototype stack operated at Plant Scholz. Laboratory testing of pilot plant and Plant Scholz CT-121 FGD gypsum, and comparisons of CT-121 FGD and phosphate gypsum engineering properties indicate the following conclusions:

- CT-121 FGD gypsum has settling, dewatering, and structural characteristics similar to, and in some instances, more favorable than phosphate gypsum, making stacking methods of waste disposal a feasible alternative. Results from geotechnical laboratory testing indicate CT-121 FGD gypsum: (1) sediments to an initial dry density greater than typical phosphate gypsums, (2) is more permeable than phosphate gypsum at equal dry densities, and (3) exhibits stress-strain-strength behavior similar to many phosphate gypsums. The pore fluid pH also has negligible effect on the engineering behavior of CT-121 FGD gypsum.

The nine-month operation of a prototype FGD gypsum stack at Plant Scholz further confirmed the feasibility of utilizing stacking for disposal of CT-121 FGD gypsum. The completed stack, approximately one-half acre (2023 m^2) and 12 feet (3.7 m) high, was generally constructed and operated as normally accomplished in the phosphate industry. Successful completion of the stack provided the following observations:

- CT-121 FGD gypsum can be stacked with a dragline using the upstream method of construction as accomplished in the phosphate industry. Based on this study, it appears the basic design and operations concepts utilized by the phosphate industry may be adopted by the utility industry for stacking FGD gypsum.

- As with some phosphate gypsum stacks, the cast CT-121 FGD gypsum dikes and slopes developed a thin drying crust which was relatively resistant to erosion from rainfall and prevented excessive dusting.

- The only significant stacking characteristic not observed in CT-121 FGD gypsum was the gradual development of some cohesion in the stack slopes from cementation. Although the self-cementation process is desirable for stacking, the absence of cementation does not preclude the use of stacking.

- As with phosphate gypsum stacks, groundwater contamination from FGD scrubber process waters is a concern. The process waters can contain concentrations of sulfate, calcium, chloride and magnesium several orders of magnitude greater than natural background levels and drinking water standards. Trace elements such as arsenic, chromium or selenium may alsobe present within the process water at levels above drinking water standards and could pose a contamination problem. Seepage from FGD gypsum stacks, therefore, must be controlled or prevented, and the surface and groundwater surrounding the stack monitored.

The effect of fly ash addition on the stacking characteristics of CT-121 FGD gypsum was briefly investigated because the potential exists for simultaneous disposal of fly ash and gypsum. The results of field observations and laboratory testing on the effect of fly ash addition on the permeability, sedimentation-consolidation and shear strength characteristics of CT-121 FGD gypsum all indicated reductions in the favorable stacking characteristics of CT-121 FGD gypsum. Additional research, however, is required to assess the stackability of fly ash-gypsum mixtures and to investigate potential methods for simultaneous disposal of fly ash and gypsum.

Waste disposal of FGD gypsum by stacking should be economically and environmentally competitive with other FGD disposal techniques, such as dry disposal/fixation processes for calcium sulfite sludges. In comparison to the other FGD wastes, FGD gypsum presents the following advantages:

- Gypsum can be easily handled by conventional construction equipment and is a more workable material than many other FGD by-products.
- Gypsum can be stacked, and can therefore be stored in a smaller area than is possible with landfilled FGD wastes.
- The operation of a gypsum stack is generally much easier and simplier than the operation of a landfill. Gypsum can be pumped to the disposal area in a slurry, therefore eliminating daily handling and transportation of wastes to a landfill site. Gypsum can be dewatered by gravity in the stacking method, eliminating the need for mechanical dewatering. Within the stacking area, gypsum quickly dewaters and consolidates to a stable material without the need for added compaction.
- Several markets exist for potential utilization of FGD gypsum, including the manufacture of wallboard and Portland cement and agricultural use as a soil amendment.
- Gypsum is a completely oxidized material, and will not exert an oxygen demand on the environment, as occurs with sulfite sludge.

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DRY ACTIVATED CHAR PROCESS FOR SIMULTANEOUS SO_2 AND NO_x REMOVAL FROM FLUE GASES

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4300 Essen 13, Federal Republic of Germany

It is important to cut as well the SO_2 as the NO_x emission from large boiler plants. A process causes especially then low costs, if SO_2 and NO_x can be removed simultaneously. The dry BF-process for flue gas desulfurization by activated char was modified for the simultanous removal of both gas components. By injecting ammonia to the flue gas in a two stage moving bed adsorber which contains activated char the nitrogen oxides are reduced catalytically to N_2 and H_2O, while the SO_2 removal efficiency is improved. This process works well also at temperatures between 120 and 160 °C, as they are usually present after the air preheater and the electrical precipitation. The temperature of the flue gas will not be changed by the process. Furthermore the dust content as well as the content of gaseous Cl- and F-components is reduced.

Results from laboratory experiments and from tests in the demonstration plant of the BF-process at Lünen are presented. The SO_2 removal efficiency is improved over 95 %, while up to 85 % of the nitrogen oxides are removed.

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DRY ACTIVATED CHAR PROCESS FOR SIMULTANEOUS SO_2 AND NO_x REMOVAL FROM FLUE GASES

Worldwide over 75 flue gas desulfurization processes have been developed, but only a limited number of them is technically applied. Although the wet processes seem to be most advanced, the main interest centers on dry processes now. They have advantages for the following reasons:
no waste water; no flue gas cooling resp. reheating; less space requirement; less energy consumption; wider variety of end products; easier removal of SO_2 and NO_x in a combined process.

In this paper, the last fact will be proved for the BF-process, in which gaseous ammonia is added to the flue gas and activated char is used as an adsorbent/catalyst for SO_2/NO_x-removal. Results from the laboratory and from the demonstration plant installed for 150.000 m^3/h (STP) show that a SO_2 removal efficiency over 95 % and a NO_x removal efficiency of about 80 % are realistic.

BACKGROUND

While the sulfur contained in the fuel is converted in the burners almost completely to gaseous sulfuric oxides, the portion of the fuel-bound nitrogen which reacts to form nitrogen oxides may vary. A further (smaller) portion originates from the oxidation of combustion air nitrogen. By combustion control measures as e.g. reduction of excess-air coefficient, multi-stage combustion, flue gas recycling etc., nitrogen oxide emissions can be reduced to a limited extent1). It is to be noted that these measures do not add up in effectiveness so that a maximum NO_x-reduction of 60 % is possible, at the present state of the art, compared to boilers without NO_x-reduction2). If nitrogen oxide emission limits as at the point of being introduced in Japan or the United States2)3), or even lower values are to be adhered to, additional measures for nitrogen oxide removal from the flue gases need to be taken. This means that, on the other hand, Denox-processes for flue gas treatment surely will not work for their own but always in combination with methods which prevent the formation of NO_x in the burners. From this point of view, a NO_x removal efficiency of 60 - 70 % in flue gas treatment is sufficient.

BF-PROCESS FOR FLUE GAS DESULFURIZATION

The principle of the Bergbau-Forschung process for the desulfurization of flue gas is the adsorptive enrichment of SO_2 of the flue gases to special activated chars made from hard coal. The adsorption stage (see Figure 1) is run in the temperature range between 100 and 150 °C without flue gas cooling. The process is suited in particular for the desulfurization of flue gases with SO_2 concentrations of up to 0,4 % by volume, corresponding to approx. 12 g SO_2/m^3 (STP), in presence of oxygen and steam, with simultaneous conversion of the separated SO_2 to adsorbed sulfuric acid:

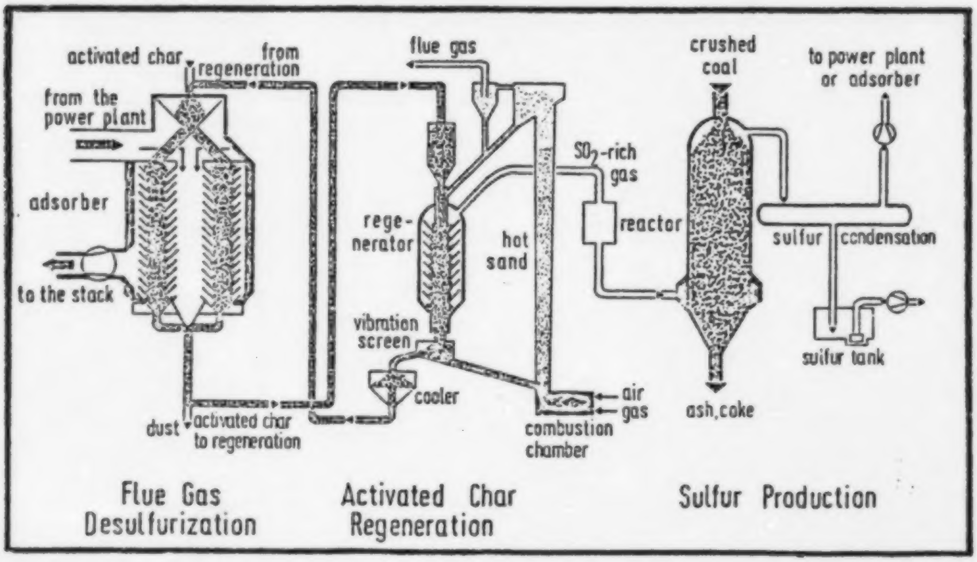

activated char
from regeneration
from the power plant
adsorber
to the stack
dust
activated char to regeneration
flue gas
regenerator
hot sand
vibration screen
cooler
combustion chamber
air
gas
SO_2-rich gas
reactor
crushed coal
ash, coke
to power plant or adsorber
sulfur condensation
sulfur tank
Flue Gas Desulfurization
Activated Char Regeneration
Sulfur Production

Figure 1 BF-process for flue gas desulfurization with activated char

$$SO_2 + \frac{1}{2} O_2 + H_2O = H_2SO_4 \qquad (1)$$

Normally, the adsorber is dimensioned for SO_2 removal efficiencies between 80 and 90 percent. The loaded activated char leaving the moving bed adsorber at the bottom is subsequently thermally regenerated, with sulfuric acid being reduced by carbon of the activated char[4]:

$$2\ H_2SO_4 + C = 2\ SO_2 + CO_2 + 2\ H_2O \qquad (2)$$

In the so produced SO_2 rich gas, the SO_2 concentration is about 30 % by volume. The regenerated activated char is cooled and again recycled to the adsorber (Figure 1). The SO_2 rich gas can be converted to sulfur by a modified Claus unit[5] by the Resox-process[6], as shown in Figure 1, or by a combination of the Resox and the Claus-process, resulting in a higher sulfur production rate from SO_2 than by the Resox-process alone. Furthermore, liquid SO_2 or sulfuric acid can be produced.

A prototype plant with a gas through-put of 150.000 m^3/h (STP), corresponding to a power plant capacity of 45 MW_{el} in the by-pass to the coal fired 350 MW_{el} block was established in the Power Plant Kellermann of STEAG AG in Lünen, FRG. It runs more than 15.000 hours successfully following the performance of this peak load power station.

Apart from SO_2, about 50 % of the chlorine and fluorine compounds of the flue gas are removed. These compounds are subsequently found in the SO_2 rich gas and converted with calcium compound in a seperate filter, installed between the adsorber and the SO_2 conversion.

BF-PROCESS FOR SIMULTANEOUS SO_2 AND NO_x REMOVAL

Due to the lower NO_x emission level aimed at in industrial countries, Bergbau-Forschung has improved the BF-process to the simultaneous removal of SO_2 and NO_x under conditions present in flue gases from coal-fired power stations after the air preheater. In this way, the boiler plant needs no further configuration.

Apart from the SO_2-binding capacity, the activated char furthermore exhibits catalytic properties for the reactions (2) and (3) between NO_x and NH_3 resp. NO_x, NH_3 and O_2 to form N_2 and H_2O:

$$6\ NO + 4\ NH_3 = 5\ N_2 + 6\ H_2O \quad (2a)$$
$$6\ NO_2 + 6\ NH_3 = 7\ N_2 + 12\ H_2O \quad (2b)$$

resp.

$$4\ NO + 4\ NH_3 + O_2 = 4\ N_2 + 6\ H_2O \quad (3a)$$
$$8\ NO_2 + 12\ NH_3 + O_2 = 10\ N_2 + 18\ H_2O \quad (3b)$$

It is supported that in the first step NH_3 and/or NO are adsorbed on the activated char before reaction (2a) takes place in several steps. In the presence of oxygen adsorbed NH_3 may be oxidized to an adsorbed NH_2 radical before the reaction (3a) with NO occurs. The same applies to NO_2 reduction.

Since SO_2 and NO_x are contained in the flue gas simultaneously, SO_2 adsorption and catalytic NO_x reduction, in case of ammonia addition, do not take place indepentently as shown in Eqns. (1) - (3). In the temperature range between 100 and 160 °C the sulfuric acid adsorbed reacts with NH_3 from the gaseous phase to form ammonium hydrogen sulfate or ammonium sulfate:

$$H_2SO_4 + NH_3 = NH_4HSO_4 \quad (4a)$$
$$NH_4HSO_4 + NH_3 = (NH_4)_2SO_4 \quad (4b)$$

These salts are deposited on the inner surface of the activated char. Since these side reactions mean increased SO_2-load and since the products, in the thermal regeneration step, are also decomposed to form SO_2, this phenomenon is made use of in industrial operation$^{7)}$.

Lab-Scale Experiments

On laboratory scale, tests were carried out in fixed-bed adsorbers. In the course of the individual tests, gas mixtures were metered to the adsorber. The gases contained 6.4 % by volume of oxygen, 9.8 % by volume of steam and variable SO_2, NO, and NH_3-concentrations. Nitrogen accounted for the balance of the gas mixture. Temperatures ranged between 120 and 180 °C; the bed length was of 1 m. At the adsorber outlet, the composition of the gas mixture was measured as a function of time. Some experimental results are shown in Figures 2 to 4.

Figure 2 shows the results of the tests carried out until steady-state outlet concentration of NO was reached (in this phase, sulfur dioxide can not yet be traced at the adsorber outlet), showing furthermore the influence of various inlet concentrations of NO and SO_2 at constant NH_3-inlet concentration and the effect of changing residence times of the gas in the adsorber on the reduction of nitrogen monoxide. The test temperature was purposely kept at the low value of 120 °C in order to investigate NO reduction under moste unfavorable conditions as they might arise in boiler plant operation. The lowest NO conversion rate is achieved with

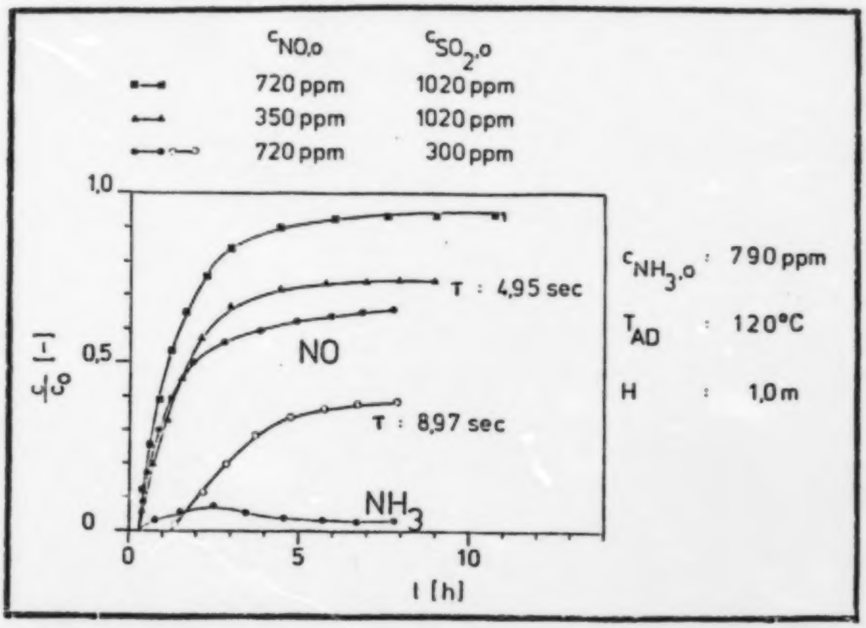


Figure 2 Break-through of simultaneous SO_2 and NO_x removal by means of NH_3 and activated char (influence of SO_2 concentration and residence time)
Adsorbent: activated char, 9 mm (Lünen)

simultaneously high SO_2 and NO concentrations (curve a). In this case, obviously too little ammonia is added (see variation of NH_3-inlet concentration in Figures 4 and 5). With lower NO-inlet concentration but unchanged concentrations of other gases, the NO conversion rate increases (compare curve b to curve a). A simplified explanation may be the following one: SO_2 reacts considerably more rapidly with ammonia than does NO, so that, with identical SO_2 and NH_3 inlet concentrations, only a similarly low quantity of not converted ammonia remains for the NO reaction. Only with excess ammonia (relative to SO_2), the NO conversion rate can be improved (curve c for reduced SO_2 inlet concentration in comparison with a and b). In this case, however, the ammonia excess is so large that ammonia is not completely converted, leaving the reactor in the proportions shown in Figure 2. This excess ammonia can be converted with NO (curve d) via increased residence time (with a correspondingly longer fixed-bed reactor). The development of ammonia concentration in the adsorber outlet in case of excess is always to be watched if ammonia proportioning is excessive: the ammonia-outlet concentration passes a maximum level after some time, where upon if decreases slowly but steadily.

This effect can also be seen in Figure 3, showing the NO break through curves for different NH_3 concentrations in the adsorber inlet for small SO_2 inlet concentrations. Ammonia, however, can be traced in the adsorber outlet only in case of largely excessive proportioning. It can be clearly seen that for the same ammonia injection the degrees of NO reduction are significantly higher than in the case of higher SO_2 inlet concentrations.

In the modified burners less NO_x is formed. Figure 4 shows break through curves for NO inlet concentrations to the reactor as they may be present in new boiler plants. It can be seen that the extent of the NO reduction can be increased to 50 percent for short residence times.
A temperatur increase to 150 °C or even higher values - without changing the other conditions - shows that in this case the NO conversion rates can be boosted.

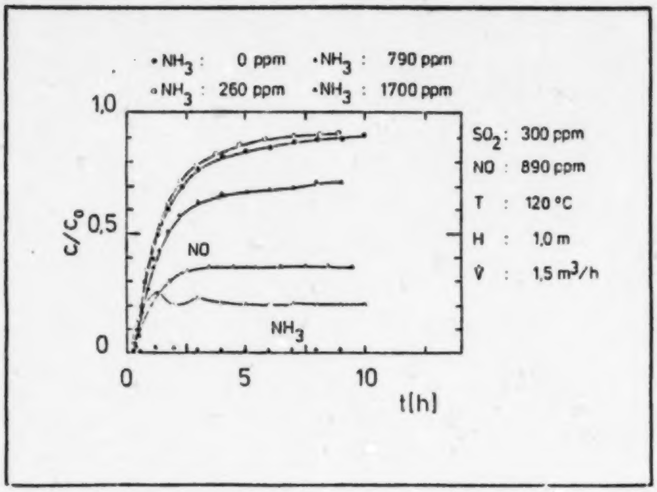


Figure 3 Influence of NH_3-inlet concentration on NO reduction and SO_2 removal at low SO_2 concentrations and high NO concentrations at T_{AD} = 120 °C
Adsorbent: activated char, 9 mm (Lünen)

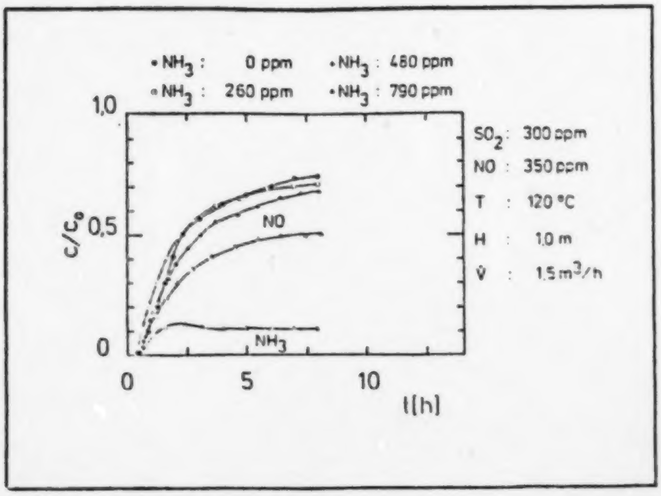
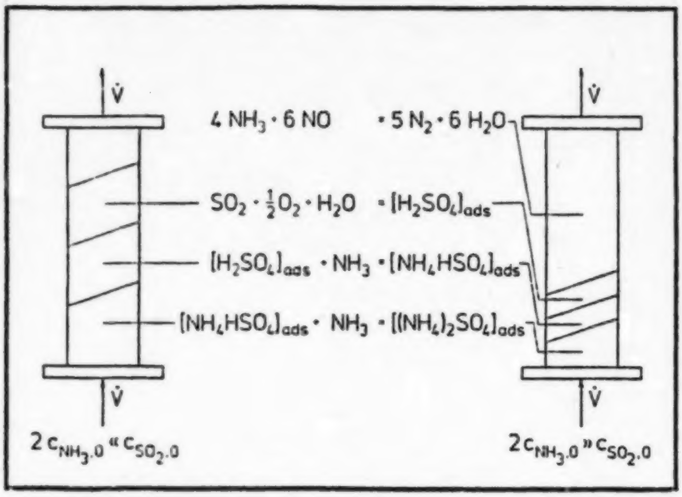


Figure 4 Influence of NH_3-inlet concentration on NO_x reduction and SO_2 removal at low SO_2 and NO concentrations at T_{AD} = 120 °C
Adsorbent: activated char, 9 mm (Lünen)

From the lab-scale experiments it can be derived that the reactions of SO_2 and NO_x with NH_3, which are parallel reactions with different rate constants, run down at different places in the reactor. This is shown for different NH_3 inlet concentrations in Figure 5. In the case of low NH_3 concentrations, ammonia reacts



$\dot{V}$
$\dot{V}$
$4\,NH_3 + 6\,NO \rightarrow 5\,N_2 + 6\,H_2O$
$SO_2 + \frac{1}{2}O_2 + H_2O \rightarrow [H_2SO_4]_{ads}$
$[H_2SO_4]_{ads} + NH_3 \rightarrow [NH_4HSO_4]_{ads}$
$[NH_4HSO_4]_{ads} + NH_3 \rightarrow [(NH_4)_2SO_4]_{ads}$
$\dot{V}$
$\dot{V}$
$2\,c_{NH_3,0} < c_{SO_2,0}$
$2\,c_{NH_3,0} > c_{SO_2,0}$

Figure 5 Pattern of the reactions of ammonia with sulfuric acid resp. nitrogen oxide in a fixed-bed reactor

mainly with adsorbed sulfuric acid, which is distributed over a relatively large part of the reactor. Ammonia reaches the parts of the reactor free from H_2SO_4 only, if the ammonia inlet concentration is higher than the SO_2 inlet concentration (right side of Figure 5). In these parts of the reactor, the NO reduction takes place. On the other hand this means that, with lower SO_2 concentrations, higher NO conversion rates are possible with the same ammonia proportioning. If highest possible NO conversion rates are required, ammonia proportioning not only upstream the single-stage adsorber (at the point of high SO_2 concentrations), but also in the zones of lower SO_2 concentrations is recommended. This is possible with a two-stage SO_2/NO-removal the simplified flow-sheet of which is shown in Figure 6. In the first stage, most of the sulfur dioxide is removed either merely adsorptively or by addition of small ammonia quantities. In this stage, the NO concentration is not reduced substantially. Upstream the second stage, ammonia is added mostly for catalytic reduction

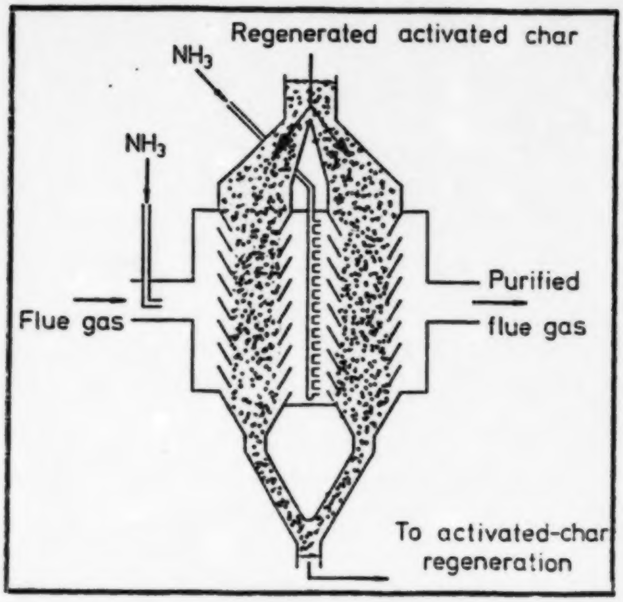


Figure 6 Dry two-stage removal of SO_2 and NO_x from flue gas according to the BF-process

of nitrogen oxides. Both stages are moving beds, which may be located in one common or in two seperated davices.

Results from the Demonstration Plant

Tests were run with the configuration shown in Figure 6 in the demonstration plant at Kellermann Power Station of STEAG AG at Lünen. Activated char as adsorbent/catalyst is used for both stages in identical quality, and also regeneration is common for activated char from both stages. The residence time of the activated char in both stages can be adjusted differently (shorter residence time in the first stage) according to the functioning mode of each stage (in the first one primarily SO_2 removal, the activated char being heavily loaded, and in the second stage primarily NO removal implying activated char being loaded moderately). In the test runs, the flue gas temperature was approx. 125 °C. The SO_2-inlet concentration was of 600 to 900 ppm. No ammonia was added to the first stage, were approx. 80 % of the SO_2 were removed. The NO_x concentration was in the range of 400-

800 ppm at the inlet and was only reduced in this process step by approx. 10 %. Upstream the second stage ammonia was proportioned. Some test results are put together in Table 1. SO_2 removal was always almost complete, while NO was reduced up to >80 percent.

Table 1 Results from the second adsorber segment at the Demonstration Plant in Lünen (SO_2-inlet concentration to the first segment: ∿1000 ppm, temperature ∿122 °C)

Date	Concentrations (ppm)				Gas residence time (s)
	SO_2 inlet	SO_2 outlet	NO_x inlet	NO_x outlet	
9.2.80	210	48	694	288	8.75
10.2.80	132	49	511	58	8.75
23.6.80	75	<20	575	116	3.75
1.7.80	65	<20	675	152	3.75
8.7.80	65	<20	475	100	4.75
14.7.80	75	<20	587	78	4.75

INDUSTRIAL APPLICATION

As the tests have shown, ammonia proportioning to a two-stage moving-bed adsorber with an activated char filling enables considerably improved SO_2 and NO removal. Compared to the classic catalysts for NO reduction used at temperatures higher than 300 °C, activated char is of considerable advantage because the inlet temperature can be lower and because SO_2 and NO_x can be removed from the flue gases simultaneously.

Compared with the original BF-process for flue gas desulfurization by SO_2 adsorption on activated char (see flow sheet in Figure 1), only little changes have to be carried out: instead of the one stage adsorber, a two stage adsorber will be used, as shown in Figure 6. Furthermore, an equipment is needed for the proportioning of ammonia.

The costs for a flue gas purification according to the BF-process with ammonia proportioning can simply and reliably be calculated by well-established data for flue gas desulfurization according to the BF-process. The ammonia proportioning does in practice not affect the capital requirement for investment since the additional costs for the ammonia supply, proportioning, and control as well as the increased costs for a two-stage instead of a single-stage adsorber are offset by reduced desorption costs. As to the utilities, ammonia supply should be allowed for in terms of approx. DM 0,001 per kWh. These additional costs are offset by reduced costs in energy demand for desorption (the activated char load, with NH_3 addition, is by 20 % higher than without NH_3 addition). Furthermore the carbon consumption of desorption is lower because sulfuric acid, in contrast to ammonium sulfate, reacts during thermal treatment with carbon. On the other hand, higher costs of the process could be justified by better NO_x suppression.

In parallel to our development, a prototype plant of similar design also using BF activated char, constructed by a Japanese company, has successfully been operating for more than one year at a Japanese power station. Scaling up is envisaged in Japan, too8)9).

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APPENDIX

Attendees

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Ahman	Stefan	POB	Vaxjo S-35187	SWEDEN		Flakt Industries, AB
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Anderson	Rick	Box 2511	Houston	TX	77001	Tennessee Gas Transmission Co.
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Bowie	Lowell C.	4413 Kings Run Drive	Cincinnati	OH	45232	Bishopric Products Company
Bowman	George A.	98 Vanadium Road	Bridgeville	PA	15017	Bowman & Associates Inc.
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Browne	William R.	135 Cumberland Road	Pittsburgh	PA	15237	Bischoff Environmental Systems
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Buchner	Gerhard H.	Getreidemarkt 9	A-1060 Vienna	AUSTRIA		Technical University of Vienna
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Carpenter	John K.	P. O. Box 149	St. Louis	MO	63166	Union Electric Company
Carrive	Francois	BP 57 91220 Bretigny-sur-Orge	Paris	FRANCE		Compagnie Generale d'Automatisme
Cartsunis	Louis P.	299 Cherry Hill Road	Parsippany	NJ	07054	Kerr-McGee Chemical Corp.
Cavanagh	Gordon	P. O. Drawer 5000	Lakeland	FL	33803	Davy McKee Corporation
Chakraverty	Chiranjit	145 Technology Park	Norcross	GA	30092	EBASCO Services, Inc.
Chambers	R. N.	8001 Daly Street	Niagara Falls, Ontario	CANADA	L56 6S2	Norton Co.
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Cline	Richard A.	10 Chatham Road	Summit	NJ	07901	MikroPul Corporation
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Coe, Jr.	E. L.	4565 Colorado Blvd.	Los Angeles	CA	90039	Joy Manufacturing Co.
Cohen	Jeffrey	901 Oak Tree Road	South Plainfield	NJ	07080	ASARCO, Inc.
Coleman	Robert	2607-H Carver Street	Durham	NC	27705	Energy & Environmental Analysis
Colley	David	8501 MoPac Boulevard	Austin	TX	78759	Radian Corporation
Collier	Clark ·	P. O. Box 173	Kansas City	MO	64141	Burns & McDonnell Eng. Co.
Conklin	Edwin R.	666 Fifth Avenue	New York	NY	10019	Pullman Power Products
Conly	John T.	520 South Post Oak	Houston	TX	77027	Bechtel Power Corporation
Cook	Charles E.	P. O. Box 473	Clifton	TX	76634	Chemical Lime Inc.
Cook	Kevin Edward	2001 Bryan Tower	Dallas	TX	75201	Texas Utilities Services Inc.

Coons	Jack D.	1800 FMC Drive West	Itasca	IL	60143	FMC Corporation
Corinthios	Hanna I. G.	880 Bay Street, 4th Floor	Toronto, Ontario	CANADA	L5A 3L7	Ontario Ministry of the Environment
Coulston	John	North Avenue at Schmale Road	Wheaton	IL	60187	Spraying Systems Co.
Cowen	Robert	85 Research Road	Hingham	MA	02043	ABCO Plastics Inc.
Cox	C. L.	601 Jefferson	Houston	TX	77002	Exxon Company, U.S.A.
Crandall	W. A.	80 Park Plaza	Newark	NJ	07101	PSE&G Research Corporation
Cress	William R.	800 Cabin Hill Drive	Greensburg	PA	15601	Allegheny Power Service Corp.
Crocker	Clarence N.	105 S. Meridian Street	Indianapolis	IN	46225	AMAX Coal Company
Crucq	Cornelis A.	Postbox 16	Vlissingen	NETHERLANDS		Royal Schelde
Cuddy	Howard K.	P. O. Box 111	Tampa	FL	33601	Tampa Electric Company
Cunic	John D.	P. O. Box 101	Florham Park	NJ	07932	Exxon Research & Engineering Co.
Cyphers	Robert	1501 Alcoa Building	Pittsburgh	PA	15219	Aluminum Company of America (Alcoa)
Czaplicki	Randi	601 Jefferson Avenue, Suite 2713	Houston	TX	77002	Lodge-Cottrell/Dresser
Czuchra	Peter A.	1800 FMC Drive West	Itasca	IL	60143	FMC Corporation
Dalton	Stuart M.	3412 Hillview Ave.	Palo Alto	CA	94304	Electric Power Research Institute
Dambra	Sal	4 Irving Place	New York	NY	10003	Consolidated Edison Company
d'Ambrosi	Faust D.	1930 Bishop Lane	Louisville	KY	40277	American Air Filter Co., Inc.
Darrough	Earl H.	High Tech. Mat'ls Div, 1020 Park Ave.	Kokomo	IN	46901	Cabot Corporation
Davidman	Sam	393 7th Avenue	New York	NY	10001	Gibbs & Hill, Inc.
Davidson	Kevin	Avon Refinery	Martinez	CA	94553	TOSCO Corp.
Davidson	Larry	100 Summer Street	Boston	MA	02110	United Engineers & Constructors
Davis	James G.	4809 Tod Avenue	East Chicago	IN	46312	Graver Energy Systems, Inc.
Davis	John R.	4901 Deramus Street	Kansas City	MO	64120	Natking & Company
Davis	Richard L.	2701 Koppers Bldg.	Pittsburgh	PA	15219	Koppers Co., Inc.
Davis	S. A.	600 Grant Street	Pittsburgh	PA	15230	U.S. Steel Corp.
Decker	Lawrence D.	S.E. Tower, Prudential Center	Boston	MA	02199	Chas. T. Main
Dedo	Damon	807 Campbell Centre II	Dallas	TX	75206	Hydro-Sonic Systems
DeHaven	Jerry E.	P. O. Box 66248	Houston	TX	77077	Peabody Process Systems, Inc.
Delis	William S.	P. O. Box 107	Union	KY	41091	Cincinnati Gas & Electric
Delleney	R. Dean	Box 9948	Austin	TX	78766	Radian Corporation
Demo	Jerry M.	6330 Hwy. 290 East	Austin	TX	78723	Texas Air Control Board
Dempsey	J. Herbert	Rt. 1, Box 423	Morrisville	NC	27709	Acurex Corp.
Dene	Charles E.	3412 Hillview Avenue, P. O. Box 10412	Palo Alto	CA	94303	Electric Power Research Institute
DePriest	William	20 South Van Buren Avenue	Barberton	OH	44303	Babcock & Wilcox Co.
Dershowitz	Mark S.	P. O. Box 1380	Houston	TX	77001	Shell Development Co.
Dickerman	James C.	P. O. Box 8650	Durham	NC	27705	Radian Corporation
Dixon	David	State House, Dept. of Env. Prot.	Augusta	ME	04333	Bureau of Air Quality Control
Doctor	Richard D.	9700 S. Cass Avenue	Argonne	IL	60439	Argonne National Laboratory
Doets	Nico A.	Centraleweg 16	Geertruidenberg	HOLLAND	4131NB	N. V. PNEM
Donahue	Bernard A.	P. O. Box 4005	Champaign	IL	61820	U.S. Army CERL
Donaldson	Temple E.	300 Liberty Street	Peoria	IL	61520	Central Illinois Light Company
Dorchester	Robert J.	P. O. Box 1700	Houston	TX	77001	Houston Lighting & Power Co.
Doty	Jane E.	576 Standard Avenue	Richmond	CA	94802	Chevron Research Company
Douglass	Barry G.	1930 Bishop Lane	Louisville	KY	40277	American Air Filter Co., Inc.
Downs	Jerry L.	464 W. Woodbury Road	Altadena	CA	91001	Meteorology Research, Inc.
Drabkin	Marvin	1820 Dolley Madison Blvd.	McLean	VA	22102	The MITRE Corporation
Drakulic	Radoje	Via G. D'Annunzio, 113	Genoa	ITALY	16121	AMN S.p.A.
Dreves	F. Carter	600 Grant St.	Pittsburgh	PA	15219	Wheelabrator-Frye, Inc.
DuBose	Richard S.	345 Courtland Street, N.E.	Atlanta	GA	30365	U.S. EPA
Duedall	Iver W.	Marine Sciences Research Center	Stony Brook	NY	11794	State University of New York
Duffy	Ronald	835 Hope St.	Stamford	CT	06907	Peabody Process Systems, Inc.
Dunkle	David M.	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Dunstone	John	1310 Indian Wood Circle	Maumee	OH	43547	American Warming & Ventilating Inc.

Duran-Lopez	Antonio	Velazquez, 132	Madrid-6	SPAIN		E.N.D.E.S.A.
Duvall	David R.	P. O. Box 849	Bloomington	IN	47402	Bloomington Crushed Stone Co., Inc.
Ebzery	Joan	Env. Reporter, 1231 25th Street N. W.	Washington	DC	20037	Bureau of National Affairs
Echter	Dana	555 17th Street	Denver	CO	80202	ARCO Coal Co.
Eckroad	Gary M.	137 Lynn Avenue, Suite 205	Ames	IA	50010	A. M. F. E. S.
Egan	Richard T.	P. O. Box 6428	Fort Myers	FL	33901	The Munters Corporation
Ekmann	James	Box 10840, Technology Center	Pittsburgh	PA	15236	U.S. Department of Energy
Elder	H. William	Energy Design & Operations	Muscle Shoals	AL	35660	Tennessee Valley Authority
Ellis	A. Jennings	Eng'g and Public Policy Dept.	Pittsburgh	PA	15213	Carnegie-Mellon University
Emerson	Robert	8420 West Dodge Road	Omaha	NB	68114	Gibbs & Hill, Inc.
Enga	Olav	1415 Johnson Drive	Madison	WI	53706	University of Wisconsin
Eriksen	Robert L.	1717 East Interstate Avenue	Bismarck	ND	58501	Basin Electric Power Cooperative
Escher	E. Dennis	1517 Woodruff Street	Pittsburgh	PA	15220	Penn Environmental Consultants, Inc.
Estcourt	V. F.	P. O. Box 3965	San Francisco	CA	94119	Bechtel Power Corporation
Evans	Brent L.	4831 North River Road	Port Allen	LA	70816	Stebbins Engineering & Manufacturing Co.
Ewing	Rita E.	550 California Street	San Francisco	CA	94104	Utah International
Farrington	Jim	1500 E. Putnam Ave., Air Corr. Div.	Old Greenwich	CT	06870	Flakt, Inc.
Favell	M. A.	555 West Hastings Street	Vancouver, B.C.	CANADA	V6B 4T6	B.C. Hydro & Power Authority
Felsvang	Karsten S.	305 Gladsaxevej	2860 Soeborg	DENMARK		Niro Atomizer, Inc.
Ferrari	Randi L.	25 Broad Street	New York	NY	10004	INFORM
Ferreira	J. Pedro	1100 Connecticut Avenue, Suite 200	Washington	DC	20036	Science Management Corporation
Fitts	Douglas R.	1500 Market St., 30th Floor	Philadelphia	PA	19106	ARCO Chemical Corp.
Flander	G. J.		Montreal, Quebec	CANADA	J3V 2Y2	Cabot Corporation
Flewelling	Frederick J.	P. O. Box 848	Copper Cliff, Ontario	CANADA	POM 1N0	C-I-L Inc.
Foley	Gerry F.	550 Kinderkamack Road	Oradell	NJ	07649	Burns & Roe, Inc.
Forbus	Robert D.	One Main Place, Suite 2700	Dallas	TX	75250	Central & South West Services
Forrest	John Alan	550 Kinderkamack Road	Oradell	NJ	07642	Burns & Roe, Inc.
Foster	Robert	P. O. Box 31000	Houston	TX	77031	Fluor Engineers & Constructors, Inc.
Fousek	Michael N.	P. O. Box 269	Prospect	KY	40059	Arelco Plastic Fabricating Co.
Fowler	Carolyn P.	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
Fox	Landon D.	400 Commerce Ave., W10B104	Knoxville	TN	37902	Tennessee Valley Authority
Fox	Raymond	9700 S. Cass Avenue	Argonne	IL	60439	Argonne National Laboratory
Frabotta	Daute	31 Inverness Ctr. Parkway	Birmingham	AL	35243	Combustion Engineering, Inc.
Fraser	Thomas L.	4809 Tod Avenue	East Chicago	IN	46312	Graver Energy Systems, Inc.
Fraunfelder	George M.	12 Holland Avenue	Peapack	NJ	07977	Komline-Sanderson Engrg. Corp.
French	William E.	Box 880	Morgantown	WV	26505	U.S. Department of Energy
Frentzen	Juergen	3201 Interstate 85 North	Charlotte	NC	28213	Babcock BSH
Frey	Steven J.	1860 Lincoln Street	Denver	CO	80295	U.S. EPA
Friggens	Gary R.	Box 880	Morgantown	WV	26505	U.S. Department of Energy
Froelich	Daniel A.	P. O. Box 173	Kansas City	MO	64141	Burns & McDonnell Eng. Co.
Fuller	Forney	1900 Veterans Boulevard	New Orleans	LA	70005	Forney Fuller & Associates
Furlong	Dale A.	3140 Chaparral	Roanoke	VA	24018	ETS, Inc.
Gage	Stephen J.	401 M Street, S.W.	Washington	DC	20460	U.S. EPA
Gaines	Joseph L.	9165 Rumsey Road	Columbia	MD	21045	Niro Atomizer, Inc.
Gallagher	Bernie	1935 West County Road B-2	Roseville	MN	55113	Minnesota Pollution Control Agency
Gardner	Arthur S.	P. O. Box 2687	Grand Junction	CO	81502	Occidental Oil Shale, Inc.
Garey	Stanford L.	10 Chatham Road	Summit	NJ	07901	MikroPul Corporation
Garvey	Burney	1300 Three Greenway Plaza East	Houston	TX	77046	Pullman-Kellogg
Garvin	W. D.	2710 Wycliff Road	Raleigh	NC	27622	Martin Marietta Aggregates Co.
Gaudette	Paul R.	6701 W 64th Street-212	Shawnee Mission	KS	66202	Research-Cottrell, Inc.
Gaur	Karan S.	P. O. Drawer 5000	Lakeland	FL	33803	Davy McKee Corporation
Gaynor	John C.	1000 E. Northwest Highway	Des Plaines	IL	60016	United States Gypsum

Gehri	Dennis C.	8900 DeSoto Avenue	Canoga Park	CA	91304	Rockwell International
Gellenbeck	Robert	215 North Front Street	Columbus	OH	43215	Columbus & Southern Ohio Electric Co.
Gellner	Douglas	10 Chatham Road	Summit	NJ	07901	MikroPul Corporation
Gentile	William N.	200 North Seventh Street	Lebanon	PA	17042	Envirotech Corp.
Gentry	H. Mickey	Three Greenway Plaza East	Houston	TX	77062	Pullman-Kellogg
Ghosh	Dipak K.	Dept. of Civil Engineering	Kanpur 208016	INDIA		Indian Institute of Technology
Giammar	Robert D.	505 King Avenue	Columbus	OH	43201	Battelle Columbus Laboratories
Gibson	Elizabeth D.	8500 Shoal Creek Blvd.	Austin	TX	78766	Radian Corporation
Gibson	John C.	1300 Park Place Bldg., 6th Ave.	Seattle	WA	98119	Chiyoda International Corporation
Giovanetti	Albert	P. O. Drawer 5000	Lakeland	FL	33803	Davy McKee Corporation
Glamser	John	P. O. Box 36444	Houston	TX	77036	DM International
Glazer	Abram R.	1900 Pennsylvania Ave., N.W., Rm.832	Washington	DC	20068	Potomac Electric Power Co.
Gleason	Robert J.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Glenn	Roland D.	50 East 41st Street	New York	NY	10017	Combustion Processes, Inc.
Goecker	Robert F.	1001 Northwest 62nd St., Suite 200	Fort Lauderdale	FL	33309	Marcona Ocean Industries, Ltd.
Goffredi	Rodger A.	P. O. Box 5406	Denver	CO	80217	Stone & Webster Engineering Corp.
Gogineni	M. Rao	1000 Prospect Hill Road	Windsor	CT	06095	Combustion Engineering, Inc.
Goldberg	Stephen M.	401 M Street, S. W.	Washington	DC	20460	U.S. EPA
Golden	Dean M.	P. O. Box 10412	Palo Alto	CA	94304	Electric Power Research Institute
Goldstrohm	Don	P. O. Box 1018	St. Johns	AZ	85936	Salt River Project
Golla	John A.	P. O. Box 35000	Houston	TX	77035	Fluor Engineers & Constructors, Inc.
Goodwin	Richard	One Penn Plaza	New York	NY	10001	Envirotech/Chemico
Grant	Richard J.	607 East Adams Street	Springfield	IL	62701	Central Illinois Public Service Co.
Graves	J. K.	7616 LBJ #550	Dallas	TX	77257	Babcock & Wilcox Co.
Greaves	Roy A.	601 Jefferson	Houston	TX	77002	Lodge-Cottrell Ltd.
Green	Bill	7134 Caenen	Shawnee Mission	KS	66216	Envirotech Corp.
Green	Clois L.	Box 472	Rockdale	TX	76567	Aluminum Company of America (Alcoa)
Greene	Jack H.	IERL, MD-60	Research Triangle Park	NC	27711	U.S. EPA
Greenstreet	W. L.	P. O. Box Y, Bldg. 9201-3 MS 2	Oak Ridge	TN	37830	Oak Ridge National Laboratory
Grey	George	9700 S. Cass Ave.	Argonne	IL	60439	Argonne National Laboratory
Grieco	Gary	2 World Trade Center	New York	NY	10048	EBASCO Services, Inc.
Grillo	Samuel A.	P. O. Box 16067	Denver	CO	80216	Mine and Smelter
Grimm	Carlton D.	40 East Broadway	Butte	MT	59701	The Montana Power Co.
Grimm	Richard Paul	4500 Cherry Creek Drive	Denver	CO	80112	Stearns-Roger Eng. Corp.
Groenewold	Gerald	Box 8156 University Station	Grand Forks	ND	58202	North Dakota Geological Survey
Gross	Robert L.	363 North Third Avenue	Des Plaines	IL	60002	Environeering, Inc.
Grothe	Lew	938 Quail Street	Denver	CO	80215	York Research Consultants
Groves	Kenneth O.	2020 Dow Center	Midland	MI	48640	The Dow Chemical Company
Gruber	Joe C.	209 E. Washington Avenue	Jackson	MI	49201	Commonwealth Associates Inc.
Gude	Klaus E.	Gladsaxevej 305, DK 2860	Soeborg	DENMARK		Niro Atomizer, Inc.
Gupta	Pat	30 South 17th Street	Philadelphia	PA	19101	United Engineers & Constructors Inc.
Gupta	Vir V.	2200 Churchill Rd., Div. APC	Springfield	IL	62706	Illinois EPA
Hackl	A.	Getreidemarkt 9	A-1060 Vienna	AUSTRIA		Technical University of Vienna
Halpern	Mark	1900 Pennsylvania Ave., N.W.	Washington	DC	20068	Potomac Electric Power Co.
Hamilton	W. Richard	835 Hope St.	Stamford	CT	06907	Peabody Process Systems, Inc.
Hanlon	Jim	1800 FMC Drive West, Air Quality Ctr.	Itasca	IL	60143	FMC Corporation
Hansen	Kurt J.	900 One Palliser Sq., 125-9th Ave.,SE	Calgary, Alberta	CANADA	T2G 0P6	Montreal Engineering Co., Ltd.
Hansen	Penelope	401 M Street, S.W., (WH-565)	Washington	DC	20460	U.S. EPA
Hansen	S. Leif	9165 Rumsey Road	Columbia	MD	21045	Niro Atomizer, Inc.
Hargrove	O. W.	P. O. Box 9948	Austin	TX	78766	Radian Corporation
Harkness	John	EES Division, Building 362	Argonne	IL	60439	Argonne National Laboratory
Harr	Bradley D.	2511 Mount, #8	Missoula	MT	59801	University of Montana
Harris, Jr.	William B.	1 River Road, Bldg. 2-132B	Schenectady	NY	12345	General Electric Company

Harrison	E. W.	601 Jefferson	Houston	TX	77002	Exxon Company, U.S.A.
Harty	David M.	8675 Sheridan Drive	Buffalo	NY	14221	Frontier Technical Associates, Inc.
Hassmann	Wilfred N.	P. O. Box 47127	Dallas	TX	75247	Gifford-Hill & Company, Inc.
Haulter	Larry		Kokomo	IN	46901	Cabot Corporation
Haverlah	Dennis W.	P. O. Box 220	Austin	TX	78767	Lower Colorado River Authority
Hayes, Jr.	Richard D.	1501 Alcoa Building	Pittsburgh	PA	15228	Aluminum Company of America (Alcoa)
Healey	Joseph	835 Hope St.	Stamford	CT	06907	Peabody Process Systems, Inc.
Heeney	John	1 Brown & Root Center	Lombard	IL	60148	Brown & Root, Inc.
Hemphill	Paul D.	601 Jefferson	Houston	TX	77005	Lodge-Cottrell/Dresser
Hennico	Alphonse	1/4 Av. Be Bois-Preau	Rveil-Malmaison	FRANCE		Institut Francais du Petroll
Henschen	Arthur	P. O. Box 66763	Houston	TX	77006	Selle Alloys & Equipment Co.
Hentges	R. A.	6105 Center Hill Road	Cincinnati	OH	45224	The Procter & Gamble Company
Henzel	David S.	11499 Chester Road	Cincinnati	OH	45246	PEDCo Environmental, Inc.
Hesketh	Howard E.	College of Engineering	Carbondale	IL	62901	Southern Illinois University (SIU-C)
Hess	H. Fred	1800 FMC Drive West	Itasca	IL	60143	FMC Corporation
Hewitt	Robert A.	2001 Bryan Tower	Dallas	TX	75201	Texas Utilities Services Inc.
Hickenbottom	Barry	Code 221	Port Hueneme	CA	93043	U.S. Navy - NESO
Hickok	Wayne	3316 West 66th Street	Minneapolis	MN	55435	Cooperative Power Association
Hicks	John	200 East Randolph	Chicago	IL	60601	FMC Corporation
Hicks	N. Dale	P. O. Box 551	Springfield	MO	65801	Springfield City Utilities
Highsmith	Charles S.	1020 9th Street	Sacramento	CA	95814	CA State Solid Waste Management Board
Hill	Fred	P. O. Box 270	Hartford	CT	06101	Northeast Utilities
Hilton	Robert G.	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Hinkley	David E.	P. O. Box 2166	Houston	TX	77001	Bechtel Power Corporation
Hoffman	David C.	650 Smithfield Street	Pittsburgh	PA	15225	Dravo Lime Company
Hoftman	Jerry	P. O. Box 173	Kansas City	MO	64141	Burns & McDonnell Eng. Co.
Holland	Michael H.	11 S. La Salle Street, Suite 2300	Chicago	IL	60603	Michael H. Holland, Attorney-at-Law
Holliday	John F.	P. O. Box 8361	S. Charleston	WV	25303	Union Carbide Corp.
Holtman	Charles B.	7927 Jones Branch Drive	McLean	VA	22102	Radian Corporation
Holtzapple	Ray A.	3535 W. 12th St.	Houston	TX	77008	Big Three Industries, Inc.
Hoover	Gerald L.	P. O. Box 1031	Elizabethton	TN	37643	Great Lakes Carbon Corp.
Horton	William M.	3700 Lake Austin Boulevard	Austin	TX	78767	Lower Colorado River Authority
Hsiang	Mary W.	Chemical Engineering Department	Austin	TX	78712	University of Texas at Austin
Hunkapillar	John H.	1020 West Park Avenue	Kokomo	IN	46901	Cabot Corporation
Hunt	Joseph	Research Center	Brackenridge	PA	15014	Allegheny Ludlum Steel Corp.
Hurst	Thomas B.	20 S. Van Buren Ave.	Barberton	OH	44203	Babcock & Wilcox Co.
Hutcheson	Royce	Tombigbee Power Plant	Leroy	AL	36548	Alabama Electric Cooperative, Inc.
Hutta	Paul J.	Somerton Road	Trevose	PA	19047	Betz Labs Inc.
Ide	A.	1300 Park Place Building	Seattle	WA	98101	Chiyoda International Corporation
Idemura	Hideo	1300 Park Place Building	Seattle	WA	98101	Chiyoda International Corporation
Illovits	Nick	2 World Trade Center	New York	NY	10048	EBASCO Services, Inc.
Innerarity	Mike	Box 2511	Houston	TX	77001	Tennessee Gas Transmission Co.
Ireland	Paul A.	Box 5888	Denver	CO	80217	Stearns-Roger Eng. Corp.
Iyer	Ramani	One Penn Plaza	New York	NY	10001	Chemico Air Pollution Control Corp.
Jackson	Craig	Route 1, Box 197-X	Bakersfield	CA	93305	Getty Oil Co.
Jain	Virenda	1500 East Putnam Avenue	Old Greenwich	CT	06870	Flakt, Inc.
Janney	Raymond B.	2385 Revere Beach Parkway	Everett	MA	02149	AVCO Research Laboratory
Jao	Yung-Wo	700 University Ave.	Toronto, Ontario	CANADA	M5G 1X6	Ontario Hydro
Jaros	Walt E.	470 Niagara Parkway	North Tonawanda	NY	14120	Metal-Cladding Inc.
Jenkins	H. Neff	P. O. Box 80609	Atlanta	GA	30366	Jenkins & Black, Inc.
Jensen	Robert M.	P. O. Box 3965	San Francisco	CA	94119	Bechtel Power Corporation
Jewell	Robert G.	1900 West Loop South, Ste. 1370	Houston	TX	77027	Pullman Power Products
Jezierski	R. P.	P. O. Box 2906	Houston	TX	77001	Shell Oil Co.

Johnson	Auldon E.	414 Nicollet Mall	Minneapolis	MN	55401	Northern States Power Company
Johnson	Carlton	835 Hope St.	Stamford	CT	06907	Peabody Process Systems, Inc.
Johnson	Todd E.	9055 Thunderhead Drive	Boulder	CO	90302	E. I. duPont De Nemours & Co., Inc.
Johnston	Ross M.	300 W. Washington Street	Chicago	IL	60606	Marblehead Lime Co.
Jones	Daniel W.	29 South La Salle Street	Chicago	IL	60187	Babcock & Wilcox Co.
Jones	J. R.	5435 Stemmons Freeway, P.O. Box 47127	Dallas	TX	75247	Gifford-Hill & Company, Inc.
Jones	Julian W.	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
Jones	N. Stuart	P. O. Box 12194	Research Triangle Park	NC	27709	Research Triangle Institute
Jones	Ronald T.	1401 N. Westshore Drive	Tampa	FL	33622	Badger America, Inc.
Jordan	Richard J.	P. O. Box 35000	Houston	TX	77035	Fluor Engineers & Constructors, Inc.
Judersleben	Peter	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Kahl	W. Keith	P. O. Box Y	Oak Ridge	TN	37830	Oak Ridge National Laboratory
Kameoka	Yohji	1300 Park Place Building	Seattle	WA	98101	Chiyoda International Corporation
Kanare	Howard M.	5420 Old Orchard Road	Skokie	IL	60077	Portland Cement Associaton
Kandall	Robert J.	433 Hackensack Avenue	Hackensack	NJ	07601	Pullman-Kellogg
Kaplan	Marilyn	2970 Maria Avenue	Northbrook	IL	60062	The McIlvaine Co.
Kaplan	Norman	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
Kasischke	Martin	807 Campbell Centre II	Dallas	TX	75206	Hydro-Sonic Systems
Kaye	Ronald	835 Hope St.	Stamford	CT	06907	Peabody Process Systems, Inc.
Keckritz	Michael L.	500 South 27th Street	Decatur	IL	62525	Illinois Power Company
Keen	Robert T.	P.O. Box 240232, Hwy. 51 & Johnston	Charlotte	NC	28224	Catalytic, Inc.
Keiner	Dan	12076 Grant Street	Thornton	CO	80241	Tri-State G & T
Keller	Jim B.	Box 5888	Denver	CO	80217	Stearns-Roger Eng. Corp.
Kelly	Brian	85 Research Road	Hingham	MA	02043	Martek Engineering Inc.
Kelly	Mary E.	3024 Pickett Road	Durham	NC	27705	Radian Corporation
Kelmers	A. D.	P. O. Box X, Bldg. 4501	Oak Ridge	TN	37830	Union Carbide Corp.
Kendle	James R.	1800 FMC Drive West	Itasca	IL	60143	FMC Corporation
Kennedy	Robert A.	P. O. Box 226331	Dallas	TX	75266	Texas Power & Light Co.
Kennedy	Tom	8506 Tybor, Room 101	Houston	TX	77074	Pennwalt-Sharples
Kent	Raymond L.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Kerr	Byron T.	363 Eastern Boulevard	Watertown	NY	13601	Stebbins International
Kesler	Richard A.	P. O. Box 16067	Denver	CO	80216	Mine and Smelter
Khan	Naseer A.	2345 Stanfield Rd.	Mississauga, Ontario	CANADA	L4Y 3Y3	Carborundum Environmental Systems
Kiff	John W.	P. O. Box 2744 Terminal Annex	Los Angeles	CA	90051	Joy Manufacturing Co. (Western Precip.)
Killion	James E.	P. O. Box 47127	Dallas	TX	75247	Gifford-Hill & Company, Inc.
King	Chris	730 North Post Oak Road, Ste. 201	Houston	TX	77024	Koch Engineering Co., Inc.
King	Lawrence P.	1562 Beeson St.	Alliance	OH	44601	Babcock & Wilcox Co.
King	Robert C.	2600 Blair Stone Road	Tallahassee	FL	32301	Dept. of Environment Regulation
Kleeburg	Ulrich	Gartnerstrasse 44	4300 Essen	W GERMANY		Gottfried Bischoff GmbH & Co., KG
Klosterbuer	James	P. O. Box 351	Cedar Rapids	IA	52406	Iowa Electric Light & Power Co.
Knape	Harold O.	1915 Sul Ross	Houston	TX	77006	Harold O. Knape & Co.
Knefelkamp	Robert L.	P. O. Box 16067	Denver	CO	80216	Mine and Smelter
Knight	R. Gordon	P. O. Box 280	Beaver	PA	15009	Michael Baker, Jr., Inc.
Koda	Hiromasa	1900 Avenue of the Stars, Suite 1165	Los Angeles	CA	90067	Kawasaki Heavy Industries Inc.
Kodras	Frank D.	Executive Campus	Cherry Hill	NJ	19083	Stone & Webster Engineering Corp.
Koehler	R. A.	1 River Road, Bldg. 4-319	Schenectady	NY	12345	General Electric Company
Komline	Russell M.	12 Holland Avenue	Peapack	NJ	07977	Komline-Sanderson Engrg. Corp.
Konvicka	Ken	P. O. Box 936	Richardson	TX	75080	Sun Production Co.
Kothari	Samir P.	P. O. Box 12744	Pensacola	FL	32575	HDR, Inc.
Kovats	Gabor	P. O. Box 226227, 4900 Singleton Blvd	Dallas	TX	75222	Glitsch, Inc.
Krasiewich	D. M.		So. Walpole	MA	02071	Bird Machine Co.
Krause	Julie	Route 1, Box 197-X	Bakersfield	CA	93308	Getty Oil Co.
Krekels	Jac. T. C.	P. O. Box 17	Sittard	HOLLAND	6130 AA	Neom B. V.

Krockta	Harry	147 E. Second Street	Mineola	NY	11501	The Ducon Company, Inc.
Kroll	Larry	11511 Katy Freeway	Houston	TX	77079	NUS Corporation
Krumme	J. Lee	Suite 200, 2555 Cumberland Parkway	Atlanta	GA	30339	Vinings Chemical Company
Kuehl	John H.	2701 S. Stoughton Road	Madison	WI	53716	Warman International, Inc.
Kuo	Wen Ling	12400 E. Imperial Highway	Norwalk	CA	90650	Bechtel Power Corporation
Kupiec	Albert R.	1517 Woodruff Street	Pittsburgh	PA	15220	Penn Environmental Consultants, Inc.
Kuykendall	Jerry W.	Building 1A, ROICC	Great Lakes	IL	60088	U.S. Navy
Kuzela	Edward V.	24 Perimeter Center East, Suite 2420	Atlanta	GA	30346	UOP Inc., Air Correction Division
Lackner	Francis J.	P. O. Box 5168	Lakeland	FL	33803	A-S-H Pump
LaMantia	Charles R.	Acorn Park	Cambridge	MA	02140	Arthur D. Little, Inc.
Lammers	Gerry	Box 400	Napierville	IL	60566	Amoco Chemicals Corp.
Lamonte	A. J.	P. O. Box 1700	Houston	TX	77036	Houston Lighting & Power Co.
Landwehr	Joseph B.	P. O. Box 173	Kansas City	MO	64141	Burns & McDonnell Eng. Co.
Lang	Edward	P. O. Box 1381	Houston	TX	77001	Stauffer Chemical Co.
Lang	Ko C.	8900 DeSoto Avenue	Canoga Park	CA	91361	Rockwell International
Langeland	A. Wes	85 Research Road	Hingham	MA	02043	ABCO Plastics Inc.
Large	David B.	2118 Milvia Street	Berkeley	CA	94704	Teknekron, Inc.
Laseke	Bernard A.	11499 Chester Road	Cincinnati	OH	45246	PEDCo Environmental, Inc.
Lavely, Jr.	Lloyd L.	Suite 212, 6701 W. 64th Street	Shawnee Mission	KS	66202	Research-Cottrell, Inc.
Lawson	P.	700 University Avenue	Toronto, Ontario	CANADA	M56 1X6	Ontario Hydro
Lawson	W. H.	P. O. Box 3	Houston	TX	77001	Brown & Root, Inc.
Lawton	John T.	108 A FIPB	Muscle Shoals	AL	35630	Tennessee Valley Authority
Layman	George O.	P. O. Box 1151	Pensacola	FL	32520	Gulf Power Company
Leeming	Peter A.	2650 Lakeshore Highway	Mississauga, Ontario	CANADA	L5J 1K4	Westroc Industries Limited
Legatski	L. Karl	1800 FMC Drive West	Itasca	IL	60143	FMC Corporation
Leitman	Jerry D.	9111 Cross Park Dr., P. O. Box 87	Knoxville	TN	37921	Carborundum Environmental Systems
Leivo	Charles C.	2408 Timberloch-C, Env. Products Div.	The Woodlands	TX	77380	Dresser Industries
Lesperance	Andre F.	196 E. Main Street	Northville	MI	48167	Dorr-Oliver Incorporated
Lewis	Clifford J.	P. O. Box 15453	Lakewood	CO	80215	National Lime Assoc.
Lillestolen	Tom	1500 E. Putnam Avenue	Old Greenwich	CT	06870	Flakt, Inc.
Lindloff	Darol S.	101 Merritt 7	Norwalk	CT	06856	UOP Inc., Air Correction Division
Lintelman	Michael R.	650 Smithfield Street	Pittsburgh	PA	15222	Dravo Lime Company
Littrell	William	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Lofdahl	Clyde A.	1301 Courtesy Road	Louisville	CO	80027	Explosive Fabricators, Inc.
Longenberger	F. H.	1860 Lincoln Street	Denver	CO	80295	U.S. EPA
Lorandos	Mike K.	10 Columbus Circle	New York	NY	10019	Power Authority of the State of NY
Lordan	David J.	1100 Milam	Houston	TX	77001	Tenneco
Lovellette	Stephen M.	4809 Tod Avenue	East Chicago	IN	46312	Graver Energy Systems, Inc.
Lovetere	David P.	P. O. Box 117A	Wheatland	WY	82201	Burns & McDonnell Eng. Co.
Love	Thomas A.	6177 Sunol Boulevard	Pleasanton	CA	94566	Kaiser Aluminum & Chemical Corp.
Lowell	Philip S.	4107 Medical Parkway, Ste. 214	Austin	TX	78756	P. S. Lowell & Company, Inc.
Luce	Kenyon J.	37 E. Huntington Drive	Arcadia	CA	91006	Carborundum Environmental Systems
Lucy	Thomas H.	Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Luedthe	Louis A.	4901 College Boulevard	Leawood	KS	66211	Niro Atomizer, Inc.
Lum	John	401 M Street, S.W., (WH-552)	Washington	DC	20460	U.S. EPA
Lunz	Ed	P. O. Box 1381	Houston	TX	77001	Stauffer Chemical
MacAleese	John E.	363 Eastern Boulevard	Watertown	NY	13601	Stebbins Engineering & Manufacturing Co.
MacArthur	Rodger	200 East Randolph Drive	Chicago	IL	60601	Amoco Chemicals Corp.
MacDonald	C. J.	555 Madison Avenue	New York	NY	10022	Thyssen-CEA
MacDonald	W. B.	P. O. Box 299	Mahone Bay, Nova Scotia	CANADA	BOJ 2E0	ABCO Plastics Inc.
MacKenzie	James S.	30 South 17th Street	Philadelphia	PA	19101	United Engineers & Constructors Inc.
Madenburg	R. S.	P. O. Box 7808, II Plaza	Boise	ID	83729	Morrison-Knudsen Co., Inc.
Maiya	P. S.	9700 South Cass Avenue	Argonne	IL	60439	Argonne National Laboratory

Majdeski	Hauk	P. O. Box 87	Knoxville	TN	37901	Carborundum Environmental Systems
Makar	John E.	1800 FMC Drive West	Itasca	IL	60143	FMC Corporation
Malki	Kal	31 Inverness Parkway	Birmingham	AL	35243	Combustion Engineering, Inc.
Manavizadeh	Ghassem B.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Mann	Earl L.	5265 Hohman Avenue	Hammond	IN	47741	Northern Indiana Public Service Co.
Manz	Oscar E.	Box 8115 University Station, UND	Grand Forks	ND	58202	Coal By-Products Utility Institute
Marcus	Herr H.	Postf. 1949/1960	5270 Gummersbach 1	W GERMANY		L & C Steinmueller GmbH
Mardirossian	Aris	1900 Pennsylvania Avenue, N. W.	Washington	DC	20068	Potomac Electric Power Co.
Martin	H. W.	7825 Park Place	Houston	TX	77087	S & B Engineers, Inc.
Martin	J. R.	P.O. Box 43030, 31 Inverness Ctr Pkwy	Birmingham	AL	35243	Combustion Engineering, Inc.
Martinez	Arthur L.	P. O. Box 2267	Albuquerque	NM	87103	Public Service Co. of New Mexico
Mattes	M. F.	20 S. Van Buren	Barberton	OH	44203	Babcock & Wilcox Co.
Maurer	Joseph T.	P. O. Box 1500	Somerville	NJ	08550	Research-Cottrell, Inc.
Maurin	Peter G.	600 Grant Street	Pittsburgh	PA	15219	Wheelabrator-Frye, Inc.
Maxwell	Michael A.	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
Maxwell, Jr.	Russell C.	P. O. Drawer 2038	Pittsburgh	PA	15230	Gulf Science & Technology
Mayfield	John R.	1310 Indian Wood Circle	Maumee	OH	43537	American Warming & Ventilating Inc.
Mazer	J. S.	1 River Road, Bldg. 4-319	Schenectady	NY	12345	General Electric Company
Mazumdar	S.	10 Chatham Road	Summit	NJ	07901	MikroPul Corporation
McAndrew	Bill	12076 Grant St.	Thornton	CO	80241	Tri-State G & T
McCormick	Charles J.	650 Smithfield St.	Pittsburgh	PA	15235	Dravo Lime Company
McCoy	Billy C.	7927 Jones Branch Drive	McLean	VA	22102	Radian Corporation
McGlamery	Gerald G.	501 CEB	Muscle Shoals	AL	35660	Tennessee Valley Authority
McGowan	Gerald F.	74 Inverness Drive East	Englewood	CO	80112	Lear Siegler, Inc.
McIlvaine	Howard	1600 Pacific Avenue	Atlantic City	NJ	08404	Atlantic City Electric
McIlvaine	Robert W.	2970 Maria Avenue	Northbrook	IL	60062	The McIlvaine Co.
McKee	Jim	P. O. Box 38	Blum	TX	76627	Round Rock Lime Company
McKenna	John D.	Suite C-103, 3140 Chaparral Dr. S. W.	Roanoke	VA	24018	Environmental Testing Services Inc.
McKennon	J. T.	Box 59	Crawford	TX	76638	Tonk Products, Inc.
McKinley	John H.	2244 Walnut Grove Ave.	Rosemead	CA	91770	Southern California Edison Co.
McNulty	Kenneth J.	850 Main Street	Wilmington	MA	01887	ABCOR, Inc.
McSweeney	Brian L.	245 Summer St., P. O. Box 2325	Boston	MA	02107	Stone & Webster Engineering Corp.
Megantz	Sam	1 River Road, Bldg. 36-120	Schenectady	NY	12345	General Electric Company
Mehta	Manan	10 Chatham Road	Summit	NJ	07901	MikroPul Corporation
Melia	Mike	11499 Chester Road	Cincinnati	OH	45246	PEDCo Environmental, Inc.
Merdes	Robert S.	P. O. Box 87	Knoxville	TN	37901	Carborundum Environmental Systems
Merlet	Herrn H.	Reuterweg 14, Postfach 3724	D-6000 Frankfurt/Main 1	W GERMANY		Lurgi Umwelt und Chemotechnik GmbH
Merrill	Richard S.	485 Clyde Avenue, MS 2-2260	Mountain View	CA	94042	Acurex Corp.
Metry	Amir A.	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Meyer	Chris E.	161 East 42nd Street	New York	NY	10017	Koch Engineering Co., Inc.
Meyler	J. A.	4565 Colorado Boulevard	Los Angeles	CA	90039	Joy Manufacturing Co.
Michener	Aubrey W.	P. O. Box 227	Waterflow	NM	87421	Public Service Co. of New Mexico
Mick	Allan	1400 S.W. Fourth	Portland	OR	97201	Boise Cascade Corp.
Micko	Richard	607 E. Adams	Springfield	IL	62701	Central Illinois Public Service Co.
Miller	Cecil E.	20 S. Van Buren Avenue	Barberton	OH	44203	Babcock & Wilcox Co.
Miller	Dick	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Miller	Robert F.	1020 West Park Ave., High Tech. Div.	Kokomo	IN	46936	Cabot Corporation
Miller	Stephen L.	176 E. Fifth Street	St. Paul	MN	55101	Burlington Industries
Minnella	Thomas J.	1500 E. Putnam Avenue	Old Greenwich	CT	06870	Flakt, Inc.
Minnier	Steven H.		Coatesville	PA	19320	Lukens Steel Company
Misra	P. R.	415 E. Paces Ferry Road, N.E.	Atlanta	GA	30305	The Taulman Company
Mitchell	Don A.	807 Campbell Centre II	Dallas	TX	75206	Hydro-Sonic Systems
Mitchell	William F.	101 Merritt 7	Norwalk	CT	06856	UOP Inc., Air Correction Division

Mobley	J. David	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
Molina	J. C.	1200 Milam St., 3500 Entex Bldg.	Houston	TX	77042	ARCO Chemical Corp.
Moll	Richard T.	955 McArns Road, Sharples Div.	Warminster	PA	18974	Pennwalt Corporation
Moore	George R.	3333 Highway 6 South	Houston	TX	77079	Shell Development Co.
Moore	Keith	8900 Desoto Avenue	Canoga Park	CA	91307	Rockwell International
Moore, Jr.	Harry G.	125 High Street	Boston	MA	01930	Cabot Corporation
Moorman	Stephen	7616 LBJ #550	Dallas	TX	75251	Babcock & Wilcox Co.
Morasky	Thomas M.	3412 Hillview Avenue, Box 10412	Palo Alto	CA	94304	Electric Power Research Institute
Morford	Robert M.	4565 Colorado Boulevard	Los Angeles	CA	90039	Joy Manufacturing Co.
Morgan	Wayne E.	1500 Meadow Lake Parkway	Overland Park	KS	66204	Black & Veatch Consulting Engineers
Moser	Robert E.	P. O. Box 3822	San Francisco	CA	94119	Brown & Root, Inc.
Mounteer	Keith J.	669 West 2nd South	Salt Lake City	UT	84110	EIMCO, PMD Corp.
Moxim	R. F.	#1 Office Park Circle	Birmingham	AL	35253	C-E Environmental Systems Division
Mulder	Willem C.	Kiggelaerstraat 15	2596TL Den Haag	NETHERLANDS		Ministry of Health & Env. Protection
Mulkey	Marcus A.	3500 Akers Road, Suite 61	Bakersfield	CA	93309	Vinings Chemical Company
Mullen	Hugh	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Mullendore	Michael G.	9165 Rumsey Road	Columbia	MD	21045	Niro Atomizer, Inc.
Murad	Fred Y.	555 Madison Avenue	New York	NY	10022	Thyssen-CEA
Muren	Edward J.	222 S. Riverside Plaza	Chicago	IL	60606	Research-Cottrell, Inc.
Murphy	Dennis	938 Quail	Denver	CO	80215	York Research Consultants
Murphy	Kenneth R.	1 River Road Bldg. 4-319	Schenectady	NY	12345	General Electric Company
Muter	Richard B.	219 White Hall, WV University	Morgantown	WV	26374	Coal Research Bureau
Nagao	Jun-ichi	8-2, Marunochi 1-chome	Chiyoda-ku, Tokyo	JAPAN		Dowa Mining Co., Ltd.
Nagasvbramanian	Mani	P. O. Box 1021R	Morristown	NJ	07960	Allied Chemical Corp.
Naumann	C. E.	300 W. Washington Street	Chicago	IL	60606	Marblehead Lime Co.
Ness	Harvey M.	Box 8213, University Station	Grand Forks	ND	58202	Grand Forks Energy Technology Center
Newhams	Thomas	835 Hope St.	Stamford	CT	06907	Peabody Process Systems, Inc.
Ng	Len F.	12 Peach Tree Hill Road	Livingston	NJ	07039	Foster Wheeler Development Corp.
Nguyen	Yen V.	800 Kipling Ave.	Toronto, Ontario	CANADA		Ontario Hydro
Nickerson	Greg	1000 Prospect Hill Road	Windsor	CT	06095	Combustion Engineering, Inc.
Nielsen	Paul T.	322 W. Duval Street	Jacksonville	FL	32202	Jacksonville Electric Authority
Nilsson	Lars-Mguar O.	505 King Avenue	Columbus	OH	43201	Battelle Columbus Laboratories
Noblett, Jr.	James G.	4107 Medical Parkway #214	Austin	TX	78756	P. S. Lowell & Company, Inc.
Noland	John W.	P. O. Box 1705	Kansas City	MO	66208	Black & Veatch Consulting Engineers
Novack	Robert	85 Research Road	Hingham	MA	02043	Martek Engineering Inc.
Novak	Dr.	Rudolfstrasse	8500 Nuernberg	W GERMANY		Grosskraftwerk Franken AG
O'Brien	Thomas F.	4565 Colorado Boulevard	Los Angeles	CA	90039	Joy Manufacturing Co.
O'Brien	Thomas J.	P. O. Box 880, Morgantown Energy Ctr.	Morgantown	WV	26505	U.S. Department of Energy
O'Dell	Charles	P. O. Box 656	Wrightsville Beach	NC	28480	LaQue Center for Corrosion Technology
O'Donnell	James J.	16200 Park Row, Industrial Park Ten	Houston	TX	77084	Pullman-Kellogg
O'Hara	Robert D.	435 Sixth Avenue	Pittsburgh	PA	15219	Duquesne Light Company
Oliver	Earl	Two Palo Alto Sq., Suite 528	Palo Alto	CA	94304	Synthetic Fuels Associates
Olson	D. G.	420 Rouser Road	Coraopolis	PA	15108	Envirotech Corp.
Ongemach	Kenneth	P. O. Box 111	Tampa	FL	33601	Tampa Electric Company
Orem	Sidney R.	700 North Fairfax Street, Ste. 304	Alexandria	VA	22314	Industrial Gas Cleaning Institute
Ostroff	Norman	835 Hope St.	Stamford	CT	06907	Peabody Process Systems, Inc.
Ottmers, Jr.	Delbert M.	8501 Mo-Pac Blvd., P. O. Box 9948	Austin	TX	78766	Radian Corporation
Oxley	Joseph H.	505 King Avenue	Columbus	OH	43201	Battelle Columbus Laboratories
Ozol	Michael A.	1450 S. Rolling Road	Baltimore	MD	21227	Martin Marietta Aggregates Co.
Padfield	Robert J.	1220 West Walnut Street	Compton	CA	90220	Cabot Corporation
Padgett	Guy V.	Consul Plaza	Pittsburgh	PA	15241	Consolidation Coal Co.
Parikh	Devendra J.	P. O. Box 3, Bldg. 91-3-NW18	Houston	TX	77001	Brown & Root, Inc.
Parikh	Dilip	10 Chatham Road	Summit	NJ	07901	MikroPul Corporation

Parker III	Charles E.	420 Rouser Road	Coraopolis	PA	15108	Envirotech Corp.
Parsons	Edward L.	200 N. 7th Street	Lebanon	PA	17042	Envirotech Corp.
Patel	Ravi	215 Central Avenue	Louisville	KY	40277	American Air Filter Co., Inc.
Patkar	Avi N.	11499 Chester Road, Chester Towers	Cincinnati	OH	45246	PEDCo Environmental, Inc.
Patton	Richard W.	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Pearson	Galen	Department 327	Golden	CO	80401	The Adolph Coors Company
Pendergraft	Lynn K.	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
Pennecke	George W.	2 Broadway	New York	NY	10004	American Electric Power
Petering	John L.	2040 Avenue "C", P. O. Box 2040	Bethlehem	PA	18001	Fuller Company
Peterson	George J.	4650 S. Pinemont St., Suite 130	Houston	TX	77041	Cabot Corporation
Petkus	Robert O.	1800 FMC Drive West	Itasca	IL	60143	FMC Corporation
Petrie	Jim	P. O. Box 227	Waterflow	NM	87421	Public Service Co. of New Mexico
Petrou	Gus	55 Batterymarch Street	Boston	MA	02110	Cochrane Steam Specialty Co.
Petti	Vincent J.	600 Grant Street, APCD	Pittsburgh	PA	15219	Wheelabrator-Frye, Inc.
Pfeffer	Steven J.	101 Merritt 7	Norwalk	CT	06856	UOP Inc., Air Correction Division
Phillips	Norman D.	P. O. Box 2040	Bethlehem	PA	18001	Fuller Company
Phung	Tan	4014 Long Beach Blvd.	Long Beach	CA	90807	SCS Engineers
Pidgeon	Mary E.	300 Lakeside Drive	Oakland	CA	94643	Kaiser Aluminum & Chemical Corp.
Pierce	Thomas	P. O. Box 880, Collins Ferry	Morgantown	WV	26505	EG&G Energy Measurements
Pietsch	David C.	400 E. Sibley Blvd.	Harvey	IL	60426	ARCO Petroleum Products Co.
Pike	Allan R.	101 Merritt 7	Norwalk	CT	06856	UOP Inc., Air Correction Division
Pineda	Marilyn M.	9822 Laporte Freeway	Houston	TX	77017	U.S. Steel Corp.
Pinson	T. Duane	1001 Northwest 62nd St., Suite 200	Fort Lauderdale	FL	33309	Marcona Ocean Industries, Ltd.
Pitman	William A.	900 Chestnut Street Tower II	Chattanooga	TN	37401	Tennessee Valley Authority
Pitts	Laura J.	222 South Riverside	Chicago	IL	60606	North American Car Corp.
Place	Barry	P. O. Box 512	Milwaukee	WI	53216	Allis-Chalmers
Plappert	James F.	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Platko	Frank E.	101 Merrit-7, P. O. Box 5440	Norwalk	CT	06856	UOP Inc., Air Correction Division
Plyler	Everett L.	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
Pohle	H. G.	125 High Street	Boston	MA	02176	Babcock & Wilcox Co.
Pope	Kenneth S.	P. O. Box 3	Houston	TX	77001	Brown & Root, Inc.
Porder, Jr.	Thomas C.	1006 N. Bowen Road	Arlington	TX	76012	PEDCo Environmental, Inc.
Power	Jon P.	One Main Place, Suite 2700	Dallas	TX	75250	Central & South West Services
Predick	Paul R.	55 East Monroe, Mechanical Division	Chicago	IL	60603	Sargent & Lundy Engineers
Princiotta	Frank T.	IERL, MD-60	Research Triangle Park	NC	27711	U.S. EPA
Provol	Steve J.	Box 2906, Room 1102 TSP	Houston	TX	77001	Shell Oil Co.
Pruske	J. E.	P. O. Box 1771	San Antonio	TX	78296	City Public Service Board
Pruzinsky	John F.	2040 Avenue "C", P. O. Box 2040	Bethlehem	PA	18101	Fuller Company
Quackenbush	Victor C.	1500 Market Street, Center Square W.	Philadelphia	PA	19102	Catalytic, Inc.
Rabb	Dave	P. O. Box 2900, Shawnee Power Plant	Paducah	KY	42001	Bechtel National, Inc.
Raben	Irwin A.	130 Sandringham South	Moraga	CA	94556	IAR Technology, Inc.
Rader	Philip C.	31 Inverness Center Parkway	Birmingham	AL	35243	Combustion Engineering, Inc.
Ralston	Gary D.	4349 E. Bayley	Wichita	KS	67218	Koch Engineering Co., Inc.
Ramil	John B.	P. O. Box 111	Tampa	FL	33601	Tampa Electric Company
Randolph	Alan	Engrg. Dept., Geology Bldg.	Tucson	AZ	85721	University of Arizona
Ranly	Hans	Hammerbacherstrasse 12-14	8520 Erlangen	W GERMANY		Kraftwerk Union AG Dept. V635
Rao	Richard	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Rasmussen	Elisabeth L.	Gladsaxevej 305	Copenhagen	DENMARK		Niro Atomizer, Inc.
Rau	Eric	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Rautzen	Robert R.	P. O. Box 1123	Dayton	OH	45401	Chemineer Agitators
Ray	William G.	235 East 42nd Street	New York	NY	10017	Pfizer Inc.
Raybourne	Richard	1201 Elm Street	Dallas	TX	75270	U.S. EPA
Redmond	James D.	3072 One Oliver Plaza	Pittsburgh	PA	15222	Climax Molybdenum Co.

Reeve	Aubrey C.	P. O. Box 400	Naperville	IL	60566	Amoco Chemicals Corp.
Reeves	Gary G.	P. O. Box 3500	Houston	TX	77035	Fluor Engineers & Constructors, Inc.
Reierstad	G. A.		So. Walpole	MA	02071	Bird Machine Co.
Reilly	John B.	555 Madison Avenue	New York	NY	10022	Thyssen-CEA
Reisdorf	Jack	4500 South Cherry Creek	Denver	CO	80217	Stearns-Roger Eng. Corp.
Renberg	W. W.	420 Rouser Road	Coraopolis	PA	15108	Envirotech/Chemico Corp.
Rhudy	Richard	P. O. Box 10412	Palo Alto	CA	94303	Electric Power Research Institute
Rice	C. H.	High Ridge Park	Stamford	CT	06904	Conoco Coal Development Co.
Richard	Ronald	P. O. Box 10	Mt. Carmel	IL	62863	Public Service Indiana, Inc.
Richardson	Larry	100 Summer Street	Boston	MA	02110	United Engineers & Constructors Inc.
Richter	Ekkegard	Franz-Fischer-Weg 61	D 4300 Essen 13	W GERMANY		Bergbau-Forschung GmbH
Riebling	William E.	800 Kinderkamack Road	Oradell	NJ	07649	Burns & Roe, Inc.
Riggs	Keith A.	12031 Kurland Drive	Houston	TX	77034	Houston Lighting & Power Co.
Riggs	William J.	Prudential Center	Boston	MA	02199	Chas. T. Main
Rochelle	Gary T.	Department of Chemical Engineering	Austin	TX	78712	University of Texas at Austin
Roe	Sheldon	Box 6428	Fort Myers	FL	33901	The Munters Corporation
Roeh	Bill	P. O. Box 66248	Houston	TX	77006	Peabody Process Systems, Inc.
Rogers	Kenneth T.	555 Madison Avenue N.W.	New York	NY	10022	Thyssen-CEA
Rohlik	Ron	Route 1, Box 197X	Bakersfield	CA	93308	Getty Oil Co.
Rolfe	Richard	22600-B Lambert St., Suite 802	El Toro	CA	92630	Mittelhauser Corp.
Rorke	Kevin C.	FMC Drive West	Itasca	IL	60143	FMC Corporation
Rosekrans	Norman A.	5780 Peachtree Dunwoody Road	Atlanta	GA	30342	Joy Manufacturing Co.
Rosenberg	Harvey S.	505 King Avenue	Columbus	OH	43201	Battelle Columbus Laboratories
Ross	R. W. (Bud)	P. O. Box 1958	Huntington	WV	25720	Huntington Alloys
Rossoff	Jerome	P. O. Box 92957	Los Angeles	CA	90009	The Aerospace Corp.
Rowland	Clyde H.	Industrial Park 10	Houston	TX	77084	Pullman-Kellogg
Rudd	Derek	P. O. Box 47837	Atlanta	GA	30362	The CADRE Corporation
Rudolph	Warren J.	P. O. Box 86	Houston	TX	77001	Lone Star Industries, Inc.
Ruggiano	Louis M.	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Ruiz	Rosa	Department of CE	Austin	TX	78712	University of Texas at Austin
Rullman	Donald H.	101 Merritt 7	Norwalk	CT	06856	UOP Inc., Air Correction Division
Runyan	R. A.	1120 Chestnut Tower 2	Chattanooga	TN	37401	Tennessee Valley Authority
Russell	Jim	P. O. Box 1700, W. A. Parish Location	Houston	TX	77001	Houston Lighting & Power Co.
Salbert	Mark	P. O. Box 96228	Houston	TX	77055	Marcona Ocean Industries, Ltd.
Saleem	Abdus	One Penn Plaza	New York	NY	10001	Envirotech/Chemico Corp.
Samish	Norman C.	P. O. Box 1380	Houston	TX	77001	Shell Development Co.-WRC
Sando	S.	1300 Park Place Building	Seattle	WA	98101	Chiyoda International Corporation
Santhanam, Jr.	Chakra J.	20 Acorn Park	Cambridge	MA	02140	Arthur D. Little, Inc.
Santos	Gerald F.	P. O. Box 35000	Houston	TX	77031	Fluor Engineers & Constructors, Inc.
Sarmiento	Otto M.	101 Merritt 7	Norwalk	CT	06856	UOP Inc., Air Correction Division
Satterlee	Dennis	5401 Gamble Drive	Minneapolis	MN	55416	HDR, Inc.
Saunders	A. E.	P. O. Box 1275	Georgetown	SC	29440	Santee Cooper/Winyah Generating Station
Scardino	Charles	1710 Goodridge Dr., P. O. Box 1303	McLean	VA	22102	Science Applications Inc.
Scarth	David	2811 Walnut Hill	Dallas	TX	75220	Peerless Manufacturing Co.
Schaezler	D. J.	7825 Park Place	Houston	TX	77087	S & B Engineers, Inc.
Schaul	Peter W.	Curtis Bldg., 6th and Walnut St.	Philadelphia	PA	19106	U.S. EPA, Region III
Scheck	Robert Wayne	Box 5888	Denver	CO	80217	Stearns-Roger Eng. Corp.
Schenk	Richard W.	400 Commerce, 595 MIB-K	Knoxville	TN	37902	Tennessee Valley Authority
Scherer	Steven M.	1900 Pennsylvania Avenue, N. W.	Washington	DC	20068	Potomac Electric Power Co.
Schlauch	Richard	567 Ridge Road	Monmouth Junction	NJ	08852	The Permutit Company, Inc.
Schneider	Mark H.	630 W. Front Street, P. O. Box 231	Wilmington	DE	19899	Delmarua Power
Schulz	David A.	230 S. Dearborn	Chicago	IL	60604	U.S. EPA
Schwartz	Richard A.	Hidden Pines Drive	Clarksburg	NJ	08510	D.R. Technology, Inc.

Schwieger	Robert G.	1221 Ave. of Americas, McGraw-Hill	New York	NY	10020	Power Magazine
Schwoyer	William	567 Ridge Road	Monmouth Junction	NJ	08852	The Permutit Company, Inc.
Scott	Bill	Cross Park Drive, P. O. Box 87	Knoxville	TN	37901	Carborundum Environmental Systems
Scroggins	James	Box 100	Granger	WY	82934	Texasgulf, Inc.
Seabrook, Jr.	B. Lawrence	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Seal	William C.	Yellow Springs Road	Devault	PA	19432	Warner Co.
Seale	William C.	3700 Lake Austin Boulevard	Austin	TX	78767	Lower Colorado River Authority
Seaward	David O.	101 Merritt 7	Norwalk	CT	06856	UOP Inc., Air Correction Division
Sedman	Charles B.	ESED, MD-13	Research Triangle Park	NC	27711	U.S. EPA
Seesee	T. A.	P. O. Box 7808	Boise	ID	83729	Morrison-Knudsen Co., Inc.
Seibel	John	P. O. Box 107	Union	KY	41091	Cincinnati Gas & Electric
Selle	Joseph B.	P. O. Box 66763	Houston	TX	77006	Selle Alloys & Equipment Co.
Semrau	Konrad	333 Ravenswood Ave.	Menlo Park	CA	94025	SRI International
Senatore	Peter J.	235 East 42nd Street	New York	NY	10017	Pfizer Inc.
Shah	N. D.	715 Horizon Drive, #380	Grand Junction	CO	81501	Multi Mineral Corp.
Sharkey	Peter S.	3100 Hamilton Boulevard	South Plainfield	NJ	07080	Victaulic Co. of America
Sharp	John A.	Chemicals Research Laboratory	McMasterville, Quebec	CANADA	J3G 1T9	Canadian Industries Ltd.
Shattuck	Donald M.	P. O. Box 5888	Denver	CO	80217	Stearns-Roger Eng. Corp.
Sherman, Jr.	Carroll H.	P. O. Box 476	Beaver	PA	15009	Morrison-Knudsen National Corp.
Sherwin	Robert M.	50 Beale Street	San Francisco	CA	94119	Bechtel National, Inc.
Shieh	Yei-Shong	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Shimizu	Taku	1118 Ichigaya Tomihisa-cho	Shinjukuku, Tokyo	JAPAN 162		Mitsubishi Heavy Industries, Ltd.
Shroff	Gajendra H.	Shady Grove Road	Gaithersburg	MD	20760	Bechtel Power Corporation
Siemak	John B.	2800 Mitchell Drive	Walnut Creek	CA	94598	Dow Chemical U.S.A.
Silence	William L.	1020 W. Park Ave.	Kokomo	IN	46901	Cabot Corporation
Simko	Alexander P.	P. O. Box 21666, MS 5188	Phoenix	AZ	85036	Arizona Public Service Co.
Sims	Max E.	40 East Broadway	Butte	MT	59701	Montana Power Company
Skamenca	D. C.	9235 Katy Freeway	Houston	TX	77024	Envirotech Corp.
Skinner, Jr.	L. P.	P. O. Box 66248	Houston	TX	77006	Process Sales Company
Skloss	Jerry L.	10620 Burnet Road	Austin	TX	78758	Pastoria Co.
Slack	A. V.	Wilson Lake Shores	Sheffield	AL	35660	SAS Corporation
Slack	Dave	5401 W. Kennedy Blvd.	Tampa	FL	33609	IU Conversion Systems, Inc.
Slakey	P.	P. O. Box 26306	Charlotte	NC	28213	Babcock BSH
Smith	Daniel B.	Ten UOP Plaza	Des Plaines	IL	60016	UOP Inc., Air Correction Division
Smith	Earl O.	1500 Meadow Lake Pkwy., P.O. Box 8405	Kansas City	MO	64114	Black & Veatch Consulting Engineers
Smith	Frank S.	Turnbridge, Hudderfield	W. Yorks	ENGLAND		Peabody Holmes, Ltd.
Smith	James R.	Box 1700	Houston	TX	77063	Houston Lighting & Power Co.
Smith	Jeffrey D.	76 S. Main Street	Akron	OH	44308	Ohio Edison Company
Smith	Michael	274 Riverside Avenue	Westport	CT	06880	Dorr-Oliver Incorporated
Smith	Michael P.	11499 Chester Road	Cincinnati	OH	45246	PEDCo Environmental, Inc.
Smith	Norman B.	Stanley Building	Muscatine	IA	52761	Stanley Consultants, Inc.
Smith	Peter V.	Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Smith	Peter V.	16200 Park Row	Houston	TX	77084	Pullman-Kellogg
Smith	Scott	P. O. Box 880	Addison	TX	75001	Metal Components
Smith	Tracey L.	P. O. Box 2180	Houston	TX	77030	Exxon Co. USA
Smithson, Jr.	G. Ray	505 King Ave.	Columbus	OH	43201	Battelle Columbus Laboratories
Smitt	Lennart	2691 McCollum Parkway	Kennesaw	GA	30144	Bahco Systems
Smock	Robert	1301 S. Grove Avenue	Barrington	IL	60010	Electric Light & Power Magazine
Smolenski	John V.	3 Executive Campus, Box 5200	Cherry Hill	NJ	08034	Stone & Webster Engineering Corp.
Smyk	Eugene B.	9700 S. Cass Avenue	Argonne	IL	60439	Argonne National Laboratory
Snapp	Jeffery C.	1000 E. Main Street	Mooresville	IN	46168	Public Service Indiana, Inc.
Snell	Jerry F.	P. O. Box 61248	Houston	TX	77208	Allied Industries

Snider	A. J.	P. O. Box 7462-A	Birmingham	AL	35223	Combustion Engineering, Inc.
Snow	Curtis M.	P. O. Box 1318	Baltimore	MD	21203	Environmental Elements Corp.
Snow	Eric	P. O. Box 576	Houston	TX	77001	Kernridge Oil Co.
Soliman	Karim S.	200 W. Monroe	Chicago	IL	60606	Fluor Power Services, Inc.
Sommer	Richard S.	226 Indian Rock Road	New Canaan	CT	06840	Sommer Industries, Inc.
Spellman	J. P.	2974 LBJ Freeway, Suite 224	Dallas	TX	75234	Envirotech Corp.
Spencer	Herbert W.	P. O. Box 2744 TA	Los Angeles	CA	90039	Joy Manufacturing Co.
Spentzen	J.	P. O. Box 26306	Charlotte	NC	28213	Babcock BSH
Sperry	H. Larry	1800 FMC Drive West, Air Quality Ctr.	Itasca	IL	60143	FMC Corporation
Spitzer	Kirk E.	101 Merritt 7	Norwalk	CT	06856	UOP Inc., Air Correction Division
Spring	Richard A.	P. O. Box 211	La Cygne	KS	66040	Kansas City Power & Light Co.
Stafurik	John F.	2020 K Street, N.W., Suite 350	Washington	DC	20006	Hagler, Bailly & Company
Stange	John R.	393 Seventh Avenue	New York	NY	10001	Gibbs & Hill, Inc.
Stauffer	Chris	Box 66248	Houston	TX	77006	Process Sales Company
Stauffer	Glenn G.	Two North Ninth Street	Allentown	PA	18101	Pennsylvania Power & Light Co.
Steele	Gene	9165 Rumsey Road	Columbia	MD	21043	Niro Atomizer, Inc.
Steeves	Harold D.	Box 299, Mahone Bay	Nova Scotia	CANADA	B0J 2C0	ABCO Plastics Inc.
Stegmann	George	4 Irving Place	New York	NY	10003	Consolidated Edison Company
Stenby	Edward W.	P. O. Box 5888	Denver	CO	80217	Stearns-Roger Eng. Corp.
Stensland	John G.	1800 FMC Drive West	Itasca	IL	60143	FMC Corporation
Stevens	Nicholas J.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Stewart	Dorothy	P. O. Box 10412	Palo Alto	CA	94303	Electric Power Research Institute
Stewart	Gerald W.	Crosby Dr., Bedford Res. Park	Bedford	MA	01730	Aerodyne Research, Inc.
Stewart	Jack F.	20 S. Van Buren Avenue	Barberton	OH	44203	Babcock & Wilcox Co.
Stewart	Lawrence L.	P. O. Drawer 5000	Lakeland	FL	33803	Davy McKee Corporation
Stewart	Merrill J.	709 Cedar Way	Oakmont	PA	15139	Chemsteel Construction Co., Inc.
Stewart	Robert F.	Box 880	Morgantown	WV	26505	U.S. Department of Energy
Stouffer	Lester E.	1 Country View Road	Malvern	PA	19355	Allen, Sherman, Hoff Company
Stowe	Donald H.	650 Smithfield Street	Pittsburgh	PA	15222	Dravo Lime Company
Straw	Harry A.	1007 Market Street	Wilmington	DE	19898	E. I. duPont de Nemours & Co., Inc.
Strong	Erwin R.	P. O. Box 400	Naperville	IL	60540	Amoco Oil Company
Stuparich	J. J.	One Penn Plaza	New York	NY	10119	Envirotech Corp.
Su	Y. P.	P. O. Box 3, Bldg. 46, Room 27	Houston	TX	77001	Brown & Root, Inc.
Swanson	Carl E.	44 Briar Ridge Road	Danbury	CT	06810	Newmont Exploration Ltd.
Swartz, Jr.	Russell L.	P. O. Box 1318	Baltimore	MD	21203	Koppers Company
Swenson	Donald O.	1500 Meadow Lake Parkway	Kansas City	MO	64114	Black & Veatch Consulting Engineers
Syler	Donald E.	1945 West Parnall Road	Jackson	MI	49201	Consumers Power Company
Symuleski	Richard A.	MC 3801, 200 E. Randolph Dr.	Chicago	IL	60601	Standard Oil (Indiana)
Takvoryan	Nurhan	1500 E. Putnam Avenue	Old Greenwich	CT	06870	Flakt, Inc.
Tan	E. Anthony	17 N. 7th Ave.	Coatesville	PA	19380	E. I. duPont de Nemours & Co., Inc.
Tarlton	Dudley	5575 Del Monte Drive	Houston	TX	77001	Bechtel National, Inc.
Taylor	James A.	1417 Lakeland Hills Blvd.	Lakeland	FL	33805	Air Pollution Consultant
Tennyson	Richard P.	P. O. Box 6428	Fort Myers	FL	33901	The Munters Corporation
Thau	Albert	10 Columbus Circle	New York	NY	10019	Power Authority of the State of NY
Thompson	Bruce A.	660 Bannock Street	Denver	CO	80206	R. W. Berk & Associates
Thompson	Carol May	8500 Shoal Creek Pkway/P. O. Box 9948	Austin	TX	78766	Radian Corporation
Tomoda	Tetsuo	8-2, Marunochi 1-chome Chiyoda-ku	Tokyo	JAPAN		Dowa Mining Co., Ltd.
Tormey, Jr.	John E.	16215 North Freeway, Suite 100	Houston	TX	77090	Zurn Industries, Inc.
Torpey	Paul J.	4 Irving Place	New York	NY	10003	Consolidated Edison Company
Trexler	Edward C.	GTN E-178	Washington	DC	20545	U.S. Department of Energy
Tummala	T. K.	P. O. Box 2130	Texas City	TX	77590	Gulf Chemical & Met. Co.
Tuttle	Lansing	3340 Peachtree Road, N. E.	Atlanta	GA	30326	APAC, Inc.
Underkofler	J. W.	P. O. Box 1975	Baltimore	MD	21203	Eastern Stainless Steel Company

Vacek	Michael G.	55 E. Monroe	Chicago	IL	60603	Sargent & Lundy Engineers
Valenta	Rudy C.	520 N. Michigan Avenue	Chicago	IL	60611	Oil-Dri Corp. of America
Van Der Brugghen	Frans W.	Utrechtseweg 310	Arnhem	NETHERLANDS		Kema
Van Meter	James A.	20-24 N. W. Fourth Street	Evansville	IN	47741	Southern Indiana Gas & Electric Company
Van Ness	Robert R.	P. O. Box 32010	Louisville	KY	50232	Louisville Gas & Electric Company
Vasan	S.	835 Hope Street	Stamford	CT	06840	Peabody Process Systems, Inc.
Vehslage	S. D.	433 Hackensack Avenue	Hackensack	NJ	07601	Pullman-Kellogg
Voelcker	Jeffrey W.	P. O. Box 635	Paoli	PA	19301	A-S-H Pump
Vogelsang	C. W.	Engineering Dept. L-3368	Wilmington	DE	19898	E. I. duPont de Nemours & Co., Inc.
Voos	Helmut	Postf. 1949/1960	5270 Gummersbach 1	W GERMANY		L & C Steinmueller GmbH
Wahl	John F.	600 Grant St., Room 2786	Pittsburgh	PA	15230	U.S. Steel Corp.
Wahl	William A.	101 Merritt 7	Norwalk	CT	06856	UOP Inc., Air Correction Division
Wakefield	David W.	1930 Bishop Lane	Louisville	KY	40277	American Air Filter Co., Inc.
Walker	Kenneth A.	Box 1318	Baltimore	MD	21203	Koppers/Enelco
Walker, Jr.	Hamilton G.	P. O. Box 1318	Baltimore	MD	21203	Koppers Company
Wallace	Anna W.	P. O. Box 12194	Research Triangle Park	NC	27709	Research Triangle Institute
Wang	Han Jang	363 N. 3rd Street	Des Plaines	IL	60016	Environeering, Inc.
Wang	Joseph K.	P. O. Box 2815	Sacramento	CA	95812	CA Air Resources Board
Wang	Kuei-Hsuing	1800 FMC Drive West, Air Quality Ctr.	Itasca	IL	60143	FMC Corporation
Wang	Shih-Chung	50 Beale Street	San Francisco	CA	94105	Bechtel National, Inc.
Ward	Robert	626 Holcomb Bridge Road	Roswell	GA	30075	Ecolaire Environmental
Watson	D. A.	50 Beale Street	San Francisco	CA	94119	Bechtel National, Inc.
Watson	Ray	4901 Broadway	San Antonio	TX	79102	Railtex, Inc.
Watts	Gene M.	Box 764	Columbia	SC	29218	South Carolina Electric & Gas Co.
Weaver	Val E.	2001 Brierhill Road	Fort Washington	MD	20022	U.S. Department of Energy
Webb, Jr.	C. A.	P. O. Box 1589	Hattiesburg	MS	39401	South MS Electric Power Assoc.
Weber	Henry C.	415 E. 52nd Street, Suite 1D	New York	NY	10022	H & W Management Science Consultants
Webster	William C.	3008 Potshop Road	Norristown	PA	19403	Webster & Assoc.
Weems	Frank	3000 Sand Hill Road	Menlo Park	CA	94022	Envirotech Corp.
Weems	William T.	210 E.P. Schock Lab	Austin	TX	78712	University of Texas at Austin
Wei	Maurice W.	1501 Alcoa Building	Pittsburgh	PA	15219	Aluminum Company of America (Alcoa)
Weir	Alexander	P. O. Box 800	Rosemead	CA	91770	Southern California Edison Co.
Weiskircher	Roy J.	600 Grant St.	Pittsburgh	PA	15230	U.S. Steel Corp.
Weiss	Lawrence H.	747 Third Avenue	New York	NY	10017	Chem Systems Inc.
Weitzer	Mandred	98, Waagner-Biro-Str.	Graz	AUSTRIA	8010	Waagner-Biro AG
Wellmon	William R.	P.O. Box 33189	Charlotte	NC	28242	Duke Power Company
Wells	Murray	P.O. Box 9948, 8500 Shoal Creek Blvd.	Austin	TX	78766	Radian Corporation
Wells	William L.	1120 Chestnut Street, Tower II	Chattanooga	TN	37401	Tennessee Valley Authority
Wentz	Ted	7th Street & Tabor Road	Philadelphia	PA	19120	Proctor & Schwartz Inc., Sub. SCM Corp.
Whipple	Jeffrey T.	550 Kinderkamack Road	Oradell	NJ	07642	Burns & Roe, Inc.
White	Charles S.	115 Gibraltar Road	Horsham	PA	19044	IU Conversion Systems, Inc.
Wickardt	Ulrich	2345 Stanfield Road	Mississauga, Ontario	CANADA	L4Y 3Y3	Carborundum Environmental Systems
Wicks	Dale A.	Stanley Building	Muscatine	IA	52761	Stanley Consultants, Inc.
Widico	Mike	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Wierschem	Robert A.	2700 One Main Place	Dallas	TX	75250	Central & South West Services
Wietbrock	Volker	Badenwerk Str. #2	7500 Karlsruhe	W GERMANY		Badenwerk
Wijdeveld	Henri			NETHERLANDS		ESMIL
Wilhelm	James H.	2394 Charros Road	Sandy	UT	84070	Codan Associates
Willett	Howard P.	835 Hope St.	Stamford	CT	06907	Peabody Process Systems, Inc.
Williams	John E.	IERL, MD-61	Research Triangle Park	NC	27711	U.S. EPA
Winter	Dale D.	P. O. Box 908	Bloomington	IN	47402	Hoosier Energy
Wist	George R.	P. O. Box 1500	Somerville	NJ	08876	Research-Cottrell, Inc.
Withrow	William	309 W. Washington, Suite 300	Chicago	IL	60606	Pollution Control Board

Wojton	Stanley	55 Public Square	Cleveland	OH	44101	Cleveland Electric Illuminating Co.
Wong	Monet	3333 Michelson Drive	Irvine	CA	92730	Fluor Engineers & Constructors, Inc.
Wood	John L.	2700 One Main Place	Dallas	TX	75250	Central & South West Services
Wright	Robert C.	P. O. Box 1000	Moberly	MO	65270	Associated Electric Coop. Inc.
Wrobel	Beth A.	1957 N. Arbogast, Apt. 1-I	Griffith	IN	46319	Northern Indiana Public Service Co.
Yamada	Jun	9-11, 1 Chome, Nihombashi Horidomecho	Tokyo	JAPAN		Kureha Chemical Industry Co., Ltd.
Yanagioka	Hiroshi	c/o 1300 Park Place Building	Seattle	WA	98101	Chiyoda Chemical Eng'g. & Const. Co.
Yarze	Joseph C.	433 Hackensack Avenue	Hackensack	NJ	07601	Pullman-Kellogg
Yeh	James	P. O. Box 10940	Pittsburgh	PA	15236	U.S. Department of Energy
Yosick	Paul	1500 E. Putnam Avenue	Old Greenwich	CT	06870	Flakt, Inc.
Zarchy	Andrew S.	P. O. Box 8	Schenectady	NY	12301	General Electric CR&D
Ziegenhorn	George J.	400 E. Sibley Boulevard	Harvey	IL	60426	ARCO Petroleum Products Co.
Zierak	Stephen J.	13231 Champion Forest Drive	Houston	TX	77069	Research-Cottrell, Inc.
Zuhlke	Steven J.	P. O. Box 173	Kansas City	MO	64141	Burns & McDonnell Eng. Co.